Design of Offshore Separation Train

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Abstract – *The main objective of this paper is to propose a thermodynamic procedure for understanding the concepts involved in designing the offshore three phase separation train and further designing the pipeline to transport the oil and gas. It has started reviewing critically the existing literature choosing the most promising thermodynamic models and systems for our task. The separator and pipeline design are predicted using these models. The viscosity, density and thermodynamic properties required in the evaluation of these selected models is determined. Emphasis is given to software tools such as Matlab and Pro II.*

Key Words: Flash calculations, Extraction of crude oil & gas, Pro II, Thermodynamics, Surface production

1. INTRODUCTION

Petroleum refers to crude oil and natural gas or simply oil and gas. These are mixtures of *hydrocarbons* which are molecules in various shapes and sizes of hydrogen and carbon atoms.

From petroleum we get numerous useful products:
i) Transportation fuels: such as gasoline, diesel fuel, jet fuel, compressed natural gas and propane
ii) Heating fuels: such as propane, liquefied petroleum gas, heating oil and natural gas
iii) Sources of electricity: such as natural gas and residual fuel oil burned to generate electricity
iv) *Petrochemicals:* from which plastics as well as some clothing, building materials and other diverse products are made.

The gas mixtures consist largely of *methane.* Natural gas usually contains the smallest hydrocarbon molecules commonly found in nature.

Ethane (two carbon, six hydrogen atoms, C2H6),
Propane (C5H8),
Butane (C4 H10), and
Natural gasoline (C5H12 to C10H22).

These four types of hydrocarbons are collectively called *natural gas liquids* (NGL) which are valuable feedstock for the petrochemical industry. When removed from the natural gas mixture, these larger heavier molecules become liquid under various combinations of increased pressure and lower temperature. *Liquefied petroleum gas* (LPG) usually refers to an NGL mix of primarily propane and butane typically stored in a liquid state under pressure. LPG (bottled gas) is the fuel in those pressurized tanks used in portable "gas". Sometimes the term LPG is referred to NGL or propane.

2. PROJECT TASK

To design a three phase offshore separator in order to produce 75000 bpd of oil from the given crude oil composition.

50% of the crude oil is coming from a nearby field in a rather steady flow, while the remaining 50% is coming from a satellite field as a sluggish flow with a typical period of 15 min and a max/min ratio of about 3.

The given crude oil composition is as follows:

Components	Mole %
Nitrogen	0.5
Carbon dioxide	1.5
Methane	50.0
Ethane	7.0
Propane	3.7
N-Butane	1.8
1-Pentane	2.2
1-Hexene	3.3
N-Hexane	2.5
1-Heptane	2.0
Pseudo-1	10.5
Pseudo-2	8.2
Pseudo-3	6.8
Total	**100.0**

The pseudo components are components with high molecular weights

Components	Molecular weight	API degree
Pseudo-1	180	0.83
Pseudo-2	250	0.86
Pseudo-3	320	0.88

The water inlet in the crude oil is 50-75% of the crude oil. Over a decay of 10 years, these numbers might double due to increased water injection to the reservoir.

The below critical pressure and temperature and other factors are taken from source R[4]

S.No	Components	Chemical Formula	Molecular weight	T_c Kelvin	P_c bar	Temp Range °c		Accentric Factor w
1	Nitrogen	N_2	28.0134	126.21	33.90			0.0400
2	Carbondioxide	CO_2	44.0098	304.13	73.75			0.2250
3	Methane	CH4	16.0426	190.56	45.99	-195.0	-183.0	0.0080
4	Ethane	C_2H_6	30.0694	305.32	48.72	-142.0	-75.0	0.0980
5	Propane	C_3H_8	44.0962	369.83	42.48	-108.0	-25.0	0.1520
6	N-Butane	C_4H_{10}	58.1230	425.12	37.96	-77.0	19.0	0.1930
7	1-Pentane	C_5H_{12}	72.1498	469.70	33.70	-50.0	58.0	0.2510
8	1-Hexene	C6H12	84.1608	504.10	32.10			0.2850
9	N-Hexane	C_6H_{14}	86.1766	507.60	30.25	-25.0	92.0	0.2960
10	1-Heptane	C_7H_{16}	100.2030	540.20	27.40	-2.0	124.0	0.3510
11	Pseudo-1	$C_{13}H_{28}$	184.3640	675.00	16.80	107.0	267.0	0.6190
12	Pseudo-2	$C_{18}H_{38}$	254.4980	747.00	12.90	172.0	352.0	0.7900
13	Pseudo-3	$C_{20}H_{42}$	282.5520	768.00	11.70	131.1	198.4	0.9070

3. INTRODUCTION

In order to design the three phase separator and further to have the efficient separation process it is important to know the thermodynamic behaviour of the components involved in the process. The initial calculations are based on the assumptions at ideal conditions.

4. CALCULATIONS AT IDEALITY

4.1 Assumptions

1) Pseudo components are one component each.
2) Given volume percentage is same as the mole percentage
3) Iso-thermal system.
4) For our initial Flash calculations, we assumed only single liquid outlet stream.

Assumption 1:

In project task, it is given three pseudo components with their molecular weights which are higher hydrocarbons, so we will distinguish each component on the basis of their molecular weight. First we have assumed it to be as alkanes and their nearest molecular weight would be chosen.

For pseudo 1 molecular weight is given as 120, so we have selected n-Octane, which has molecular weight of 114. Similarly component pseudo 2 have the molecular weight of 180 and we assumed it as tri-decane which has the molecular weight of 184. For third one we assumed it as Octa decane, which has the molecular weight of 254 which is close enough to 250.

Given Components	Mol.Wt	Selected Components	Mol.Wt
Pseudo 1	120	n-Octane	114
Pseudo 2	180	n-tri-decane	184
Pseudo 3	250	n-Octa-decane	254

Assumption 2:

We have assumed that given volume percentage is same as the mole percentage

Assumption 3:

We have assumed that temperature will not vary during the separation process i.e. the system will be maintained at constant temperature.

Assumption 4:

Flash calculations: In our flash calculations at ideality, we assumed one liquid outlet stream, so let us consider for our system that 'F' moles of feed is entering with an overall composition (z_i). Let 'L' be the moles of liquid at the outlet with composition of (X_i) and 'V' be the moles of vapour at the outlet with an overall composition of (Y_i).

The overall mass & component balance on the system would yield the following equations

$$F = L + V \qquad \text{--------------------(1)}$$

$$F (z_i) = L (X_i) + V (Y_i) \qquad \text{------------------- (2)}$$

From eq. (1) we have   L = F –V

We know that equilibrium constant is given by, $K_i = \dfrac{Y_i}{X_i}$

(*Procedure to calculate Ki is explained in next step*)
So combining the equations (1) & (2) with the above relations we have,

$$Y_i = \frac{z_i}{(1 - \dfrac{V}{F})\dfrac{1}{Ki} + \dfrac{V}{F}} \qquad \text{--------- (3)}$$

So once we calculate Mole-fractions of components in vapour phase (Y_i), we can then find (X_i) from the equilibrium constant relation given by

$$X_i = \frac{Y_i}{K_i} \qquad \text{------------------ (4)}$$

As we know that sum of mole-fractions will always be unity, so (V/F) can found by assuming a value satisfying the following equation

$$\sum X_i = \sum Y_i = 1$$

(*Note*: *Since these mole-fractions (Xi & Yi) are obtained by making assumptions at Ideality, so they are used only as the initial guess values in our real case*)

4.2 Calculation of equilibrium constant

Definition of Raoult's law: The partial pressure (P_i) of a component (i) in vapour phase is at an adjusted equilibrium proportional to the mole fraction (X_i) in the liquid phase. For an ideal mixture, Raoult's law states that

$$Y_i P = P_i = P^{sat} X_i$$

For an ideal mixture, $K_i = \dfrac{P^{sat}}{P}$

Therefore $K_i = \dfrac{P^{sat}}{P} = \dfrac{Y_i}{X_i}$ ------------------ (5)

K_i depends on the pressure (P), temperature (T) and the composition of mixture.
and the saturated vapour pressure (P^{sat}) depends on equilibrium temperature.

The saturated vapour pressure (P^{sat}) can be calculated using '*Antoine Equation*' which is given as follows:

$$\text{Log } (P_i{}^{sat}) = A_i - \frac{B_i}{(T + C_i)} \quad \text{--------------- (6)}$$

Where,

    $P_i{}^{sat}$ = Saturated vapour pressure of component (i) in mm Hg

    A_i, B_i & C_i = Antoine constants of component (i)

    T = System temperature in Kelvin

Once we get the values of P^{sat}, we can now calculate K_i using eq. (5)

5. BUBBLE POINT AND DEW POINT CALCULATIONS

By definition when a liquid is heated to a certain temperature, the point at which the first bubble forms is known as the 'Bubble Point'. At this point, the temperature and pressure would be T = T $_{bubble}$ &   P = P $_{bubble.}$

$$P_{bubble} = \sum P_i^{sat} X_i$$

Where,

    P $_{bubble}$ = pressure at which the first bubble forms

    $P_i{}^{sat}$ = Saturated Vapour Pressure

    X_i = Liquid mole fraction

      (At this point $X_i = Z_i$)

Similarly when a vapour is condensed at a certain temperature and pressure, the point at which the first liquid droplet is formed is known as the 'Dew Point'. At this point the temperature and pressure are T = T $_{dew}$ & P = P $_{dew}$.

$$P_{dew} = \frac{1}{\sum \left(\dfrac{Y_i}{P_i^{sat}} \right)}$$

Where,

    P $_{dew}$ = pressure at which the first bubble forms

    $P_i{}^{sat}$ = Saturated Vapour Pressure

    Y_i = Vapour mole fraction

      (At this point $Y_i = Z_i$)

6. CALCULATIONS AT REALITY

In contrast to our assumption at ideality, the given hydro carbon compounds will be present in some quantity in all the vapour, oil & water phases. As we know that oil & water are immiscible liquids, so we have two separate liquid streams as shown in figure 1.

6.1 Flash calculations

So let us consider for our real system that 'F' moles of feed is entering with an overall composition (z_i). Let 'L^1' be the moles of oil at the outlet with composition ($X_i{}^1$), 'L^2' be the moles of water with composition ($X_i{}^2$) and 'V' be the moles of vapour with an overall composition of (Y_i).
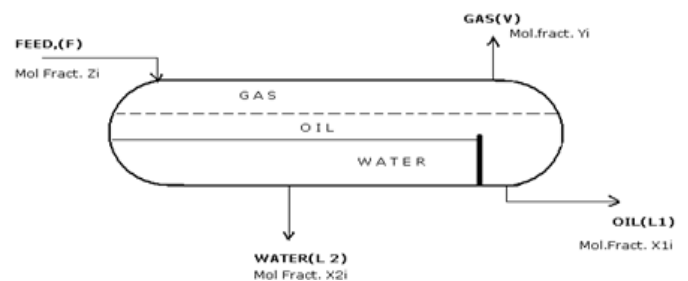

Fig -1. Separation of Oil, Gas and Water

Making a component balance on the whole system gives us,

F (Z_i)= L^1 ($X_i{}^L$) + L^2 ($X_i{}^W$) + V (Y_i)   ---------------- I

 Known terms : F & Z_i &   Unknown terms : L^1, L^2, V, $X_i{}^L$, $X_i{}^W$ & Y_i

6.2 Concept of splits

We have one equation with many unknown variables so we need to construct few more equations, so let us consider the concept of splits.

a) Vapour - liquid split (ψ): It is the ratio of vapour to feed

    Ψ = V/F $\rightarrow$ V = Ψ F   ----------------(9)

b) Oil – water split (ε): It is the ratio of Oil to the total liquid (oil+ water)

    ε = $L^1/(L^1 + L^2)$ $\rightarrow L^1 = (L^1 + L^2)\,\varepsilon$ ----------- (i)
$L^1 + L^2$= Total Liquid = Total Flow – vapour flow

$L^1 + L^2$= F –V = F - Ψ F =F (1- Ψ)

Substituting in eq. (i) L^1 = F (1- Ψ) ε ------- (10)

Similarly we have L^2 = F (1- Ψ) (1- ε) --------(11)

6.3 Conditions at equilibrium

Fugacity is the tendency to vaporize, it is a corrected pressure function 'f' and the ratio of fugacity to pressure is defined as the fugacity coefficient (Φ).

For a multi component system

$$(\Phi) = f_i / (Pz_i)$$

Where zi is the mole-fraction of component 'i' .Since all the systems behave as ideal gases at very low pressures

$$(\Phi) \rightarrow 1 \quad when \quad P \rightarrow 0$$

The departure of fugacity coefficients from unity is therefore, a measure of non-ideality of the system.

At equilibrium, fugacity of a component in vapour phase will be equal to that in a liquid phase.

$$f_i^V = f_i^L$$

So by definition we have, Fugacity, f = P Φ

$$Y_i P \phi_i^v = X_i^L P \phi_i^L$$

$$Y_i \phi_i^v = X_i^L \phi_i^L \qquad \text{---------------- (ii)}$$

$$Y_i = X_i^L \left(\phi_i^L \Big/ \phi_i^v \right) \qquad \text{-------------- (12)}$$

$$f_i^V = f_i^W$$

$$Y_i \phi_i^v = X_i^W \phi_i^W \qquad \text{-------------- (iii)}$$

From (ii) & (iii) we have

$$Y_i \phi_i^v = X_i^L \phi_i^L = X_i^W \phi_i^W$$

$$X_i^W = \left(\frac{\phi_i^L}{\phi_i^W} \right) X_i^L \qquad \text{-------------- (13)}$$

Using all these eq. from (9) to (13) and substituting in eq. (I) yields

F (Z_i) = F {Ψ (Φ_i L/ Φ_i V) X_i^L + (1- Ψ) ε X_i^L + (1- Ψ) (1-ε) (Φ_i L/ Φ_i W) X_i^L}

Rearranging and simplifying the above equation gives us the final mole-fractions of components in Liquid, Water and Vapour Phases as follows:

$$X_i^L = \frac{z_i}{\psi \left(\frac{\phi_i^L}{\phi_i^v} \right) + (1-\psi)\varepsilon + (1-\psi)(1-\varepsilon) \left(\frac{\phi_i^L}{\phi_i^W} \right)} \qquad \text{--------- (14)}$$

$$X_i^W = \frac{z_i \left(\frac{\phi_i^L}{\phi_i^W} \right)}{\psi \left(\frac{\phi_i^L}{\phi_i^v} \right) + (1-\psi)\varepsilon + (1-\psi)(1-\varepsilon) \left(\frac{\phi_i^L}{\phi_i^W} \right)} \qquad \text{--------- (15)}$$

$$Y_i = \frac{z_i \left(\frac{\phi_i^L}{\phi_i^v} \right)}{\psi \left(\frac{\phi_i^L}{\phi_i^v} \right) + (1-\psi)\varepsilon + (1-\psi)(1-\varepsilon) \left(\frac{\phi_i^L}{\phi_i^W} \right)} \qquad \text{--------- (16)}$$

6.4 Split calculations

In order to calculate the mole-fractions from equations (14,15 &16), we have to know the splits (Ψ & ε) and the fugacities of components in vapour, liquid & water phases ($\Phi_i{}^V, \Phi_i{}^L, \Phi_i{}^W$).

We have many variables but very few equations, so we have to construct some extra equations.

We know that sum of mole-fractions is unity, i.e.,

$$\Sigma X_i^L = 1, \Sigma X_i^W = 1 \text{ and } \Sigma Y_i = 1$$

So for obtaining splits we can construct & solve the equations given below (Henley-Rosen Algorithm)

$$\Sigma X_i^L - Y_i = 0 \qquad \text{--------------- (17)}$$
$$\text{and } \Sigma X_i^L - X_i^W = 0 \qquad \text{--------------- (18)}$$

By solving the eq. (17 &18), we obtain Vapour-liquid and Oil-Water splits (Ψ & ε).

6.5 Peng-Robinson equation of state

The generalized Peng-Robinson equation of state which is commonly used to represent the hydro carbons and inorganic gases (such as nitrogen, oxygen & hydrogen sulphide) is given by the following equation

$$P = \frac{RT}{\underline{v} - b} - \frac{a}{\underline{v}(\underline{v} + b) + b(\underline{v} - b)} \qquad \text{---------- (19)}$$

Where,

$$a = 0.45724 \frac{R^2 T_C^2}{P_C} \alpha \qquad \text{-------- (19.1)}$$

$$b = 0.07780 \frac{RT_C}{P_C} \qquad \text{-------- (19.2)}$$

$$\sqrt{\alpha} = 1 + k \left(1 - \sqrt{\frac{T}{T_C}} \right) \qquad \text{-------- (19.3)}$$

$$k = 0.37464 + 1.54226\omega - 0.26992\omega^2$$

All Equations of states can be expressed in cubic form known as 'cubic equation of state'. All these cubic equations are approximate; generally they provide a reasonable description of PVT behaviour in both the vapour & liquid regions for hydro carbons. In our case we have chosen Peng-Robinson Equation of state and it can be expressed in '*cubic form*' as follows:

$$Z^3 + (B-1)Z^2 + (A - 3B^2 - 2B)Z + (B^3 + B^2 - AB) = 0 \qquad \text{---------- (20)}$$

Where, $\quad Z = \frac{P\underline{v}}{RT}$

$$A = \frac{aP}{R^2T^2}$$
$$B = \frac{bP}{RT}$$

6.6 Mixing rules

All Equations of State are applied to multi-components systems by employing mixing rules to determine their parameters for the mixtures. The parameters of Equation of State are considered to represent the attractive and repulsive forces between the molecules. Hence the mixing rules should describe the prevailing forces between the molecules of different substances forming the mixture.

The attractive force between the molecules ' i ' and ' j ' represented in Equation of State by parameter a_{ij} which is of energy nature is given by the following equation

$$a = \sum_i \sum_j X_i X_j a_{ij} \qquad \text{----------------- (21)}$$

Where,

$$a_{ij} = \sqrt{a_{ii} a_{jj}}(1 - k_{ij})$$

'K_{ij}' is an additional parameter to express the attractive term between pairs of non-similar molecules known as '*Binary interaction parameter*'. For hydro carbons 'K_{ij}' is assumed to be zero.

The repulsive force between the molecules 'i' and 'j' represented in Equation of State by parameter b_{ij} which has the characteristic of volume is given by the following equation

$$b = \sum_i X_i b_i \qquad \text{--------------- (22)}$$

The above mixing rules known as the 'Vander Waals Mixing Rules' treat all the components similarly hence they are also referred to as 'Random Mixing Rules'.

6.7 Fugacities

We can obtain the fugacities from the equation given below derived from Peng- Robinson equation.

$$\ln \phi_i = \frac{b_i}{b}(Z-1) - \ln\left(Z - \frac{bP}{RT}\right) - \frac{a}{2\sqrt{2}bRT}\left(\frac{2\sum Y_i a_{ij}}{a} - \frac{b_i}{b}\right)\ln\left(\frac{Z + (1+\sqrt{2})bP/RT}{Z + (1-\sqrt{2})bP/RT}\right)$$

$$\text{-------------(23)}$$

To calculate the fugacity of each species in vapour, oil and water phases using the above equation at the specified values of temperature (T) , pressure (P) and mole-fractions of components in all the three phases(X_i^L, X_i^w, Y_i), the following procedure is used.

Obtain the parameters a_{ii} and b_i for each component of mixture by fitting the pure component data using the following correlations.

$$a_{ii} = 0.45724\frac{R^2 T_{Ci}^2}{P_{Ci}}\alpha_i$$

$$b_i = 0.07780\frac{RT_{Ci}}{P_{Ci}}$$

$$\sqrt{\alpha_i} = 1 + k_i\left(1 - \sqrt{\frac{T}{T_{Ci}}}\right)$$

$$k_i = 0.37464 + 1.54226\omega_i - 0.26992\omega_i^2$$

Compute 'a' and 'b' for mixtures using the following mixing rules (K_{ij} for hydro- carbons is assumed to be zero)

$$a = \sum_i \sum_j X_i X_j a_{ij}$$

Where ($a_{ij} = \sqrt{a_{ii} a_{jj}}(1 - k_{ij})$)

$$b = \sum_i X_i b_i$$

3) Then solve the Peng Robinson cubic equation for the vapour and liquid compressibility.

4) Once we obtain these values of compressabilities (Z^v, Z^L, Z^w) from the

step- 3, we can then calculate the fugacities for each species using the equation (23) for i= 1 to n (where n= number of components)

7. PROCEDURE FOR VOLUME CALCULATIONS

1 Try to obtain the parameters a_{ii} and b_i of Equation of State for the individual components from the equations (19.1 &19.2) using the pure component data.

2) Once we obtain these values try compute 'a' and 'b' for mixtures using the Vander Waals mixing rule.

3) When the values of 'a' and 'b' for mixtures have been obtained then calculate the compressibility factors for the vapour and liquid phases using Peng Robinson Cubic equation (refer to eq.(20))
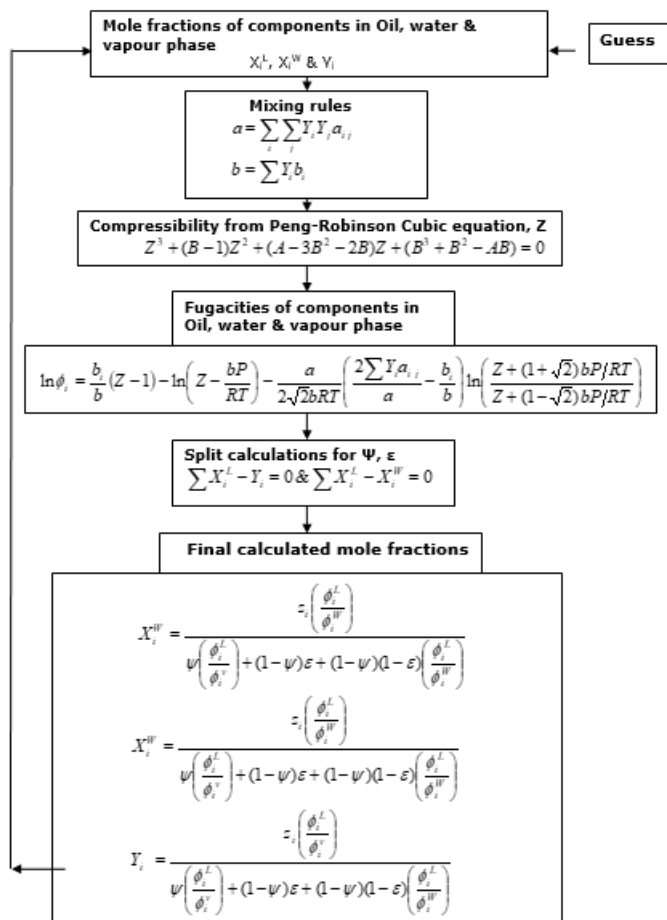
Fig -2 Algorithm for overall thermodynamic calculations

4) As we know that compressibility is defined by the following relation,

$$Z = \frac{P\underline{v}}{RT} \qquad \text{------------ (24)}$$

Molar Volume $\underline{v} = \frac{ZRT}{P}$

5) Or, we can even calculate the volume from the following equation of state once we get the values of parameters a & b for mixtures.

$$P = \frac{RT}{\underline{v}-b} - \frac{a}{\underline{v}(\underline{v}+b)+b(\underline{v}-b)}$$

8. DENSITY CALCULATIONS

For gas mixtures: From the Equilibrium constant values, it is quite evident that first four components in HP and LP Separator would go to gas phase. We can then calculate the density for the gas mixture using the following formula.

$$\rho = \rho_M * M \qquad \text{------------ (25)}$$

Where,

$\rho_M = (1/V) = P / Z_v RT$ (Molar density)

Z_v = Compressibility factor of gas mixture

R = Universal gas constant

M = Σ (Yi*Mi) Gas molecular weight

P & T are Pressure and Temperature

For liquid mixtures: The Density for the liquid mixture is obtained by the following equation.

$$\rho = 1 / \Sigma (wi/ \rho i) \qquad \text{------------ (26)}$$

Where,

wi = mass fraction

ρi = density of individual components

In our case we have found out that density of liquid for both separators to be,

HP - 766.58kg/m³   &   LP - 767.31 kg/m³

9. VISCOSITY CALCULATIONS

Gas viscosity: The gas viscosity generally increases with pressure. The increase of temperature decreases the liquid viscosity, whereas it increases the gas viscosity at low and moderate pressures. At high pressures the gas viscosity behaviour approaches that of liquid. The following correlation [R 1] can be used to estimate the gas viscosity over a temperature range of 560°R (311.1K) to 800°R (444.1K) up to 8000psia (bars).

$$\mu_g = 10^{-4} a \exp\left[b\left(\frac{\ell_g}{62.43} \right)^c \right] \qquad \text{-------- (27)}$$

Where

$$a = \frac{(9.379 + 0.0160M)T^{1.5}}{(209.2 + 19.26M + T)}$$

$$b = 3.448 + 0.01009M + \frac{986.4}{T}$$

$$c = 2.4 - 0.26$$

M = Gas molecular weight;

ℓ_g = density of gas (lbm/ft3)

Parameters	HP Separator	LP Separator
M	17,187	17,372
ℓ_g (lb/ft3)	0,8708	0,4464
T	563,4°R(313 K)	572,4°R(318 K)
a	116,9801	118,4822
b	5,3722	5,34655
c	1,3255	1,3306
μ_g	0,01191 cp	0,01193 cp

Liquid viscosity: The Beggs and Robinson equation[R 4] relating viscosity, gravity and temperature can be used to estimate the Oil viscosity which is given as follows

$$\mu_{oil} = 10^A -1 \quad (\mu_{oil} = \text{Oil viscosity in centi-poise, cp})$$

$$A = B (T)^{-1.163}$$

$B = 10^C$
$C = 3.0324 - 0.02023*G$
G = oil gravity (º API) ;
T = Temperature (ºF)

The data set from which this relationship was obtained included a range of between 16^0 & 58^0 API and 70ºF to 295ºF

Separator	T(ºF)	G	C	B	A	μ_{oil} (cp)
HP	103.73	53.0843	1.9585	90.8875	0.41115	1.5772
LP	112.73	52.909	1.9620	91.6327	0.37629	1.3784

10. THERMAL CONDUCTIVITY

For liquids: The Weber equation[R 4] can be used to make a rough estimation of thermal conductivity of organic liquids.

$$k_T = 3.56*10^{-5} C_p (\rho^4/M)^{1/3} \quad ------ (28)$$

Where,

ρ = Density of liquid mixture Kg/m³

C_p = Specific heat capacity KJ/Kg ºC

M = Molecular mass

Parameters	HP Separator	LP Separator
Liquid density, ρ(Kg/m³)	766.58	767.31
Specific heat, C_p(KJ/Kg ºC)	2.1352	2.1771
Molecular Mass, M	151.09	162.37
Thermal conductivity, k_T (W/mºK)	0.10013	0.09979

11. FLOW RATE CALCULATIONS

As we were given the task of producing 75,000 bpd of oil, so that should be the oil coming out of LP Separator as shown **in figure.4.3** (i.e the stream O2)

Volumetric flow rate = 75000 x 0.15899/ 24 =496.84 m³/ hr
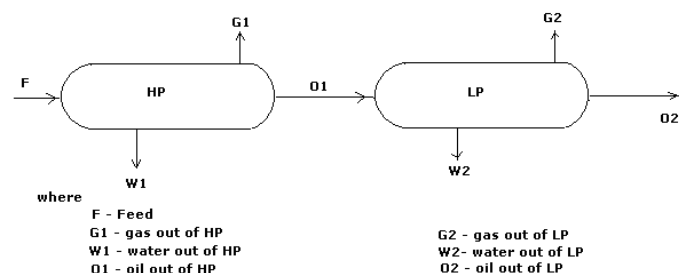Mass flow rate of 'O2' = vol. flow rate x density of oil (in LP) =381236.650 kg / hr

Fig 3. Flow rates of Oil, Gas and Water streams

For LP Separator:

Since we don't know the flow rates of any other stream other than stream 'O2', let us start our mass balance initially with LP Separator.

$$O1 = G2 + W2 + O2 \quad ------------- (a)$$

And from our program output, we know that for LP, $\Psi = 0.060$ & $\varepsilon = 0.8019$

Further we have, $\psi = \dfrac{vapour}{feed} = \dfrac{G2}{O1}$

i. e $G2 = \Psi(O1)$ ------------ (b)

And $\varepsilon = \dfrac{oil}{total\,liquid} = \dfrac{O2}{O2+W2}$

$\dfrac{O2}{O2+W2} = 0.8019 = \dfrac{381236.650}{381236.650+W2}$

W2 = 94180.047 kg/hr ------------ (c)

Sub eq. (b) in eq. (a) we have

$$O1 = \Psi(O1) + W2 + O2$$
$$O1 = \dfrac{W2+O2}{(1-\psi)} = (94180.047+381236.650)/(1-0.060)$$

O1 = 505762.4436 Kg/hr ------------- (d)

Sub above eq. (d) in eq. (b) we have G2 = 30345.7466 kg/hr ---------- (e)

For HP Separator:

Mass balance on HP Separator yields
$$F = G1 + W1 + O1 \quad -------------- (f)$$
And from our calculations, we know that for LP, $\Psi = 0.3183$ & $\varepsilon = 0.3794$

Further we have, $\psi = \dfrac{vapour}{feed} = \dfrac{G1}{F}$

i. e $G1 = \Psi(F)$ -------------- (g)

And $\varepsilon = \dfrac{oil}{total\,liquid} = \dfrac{O1}{O1+W1}$

$\dfrac{O1}{O1+W1} = 0.3794 = \dfrac{505762.4436}{505762.4436+W1}$

W1 = 827296.1848 Kg/hr -------------- (h)

Sub eq.(g) in eq.(f) we have, $F = \Psi(F) + W1 + O1$

$$F = \dfrac{W1+O1}{(1-\psi)} = \dfrac{827296.1848+505762.4436}{(1-0.3183)}$$
F = 1955491.607 Kg/hr --------------- (i)

Substituting eq. (i) in eq. (g) we have G1 =622432.9785 kg/hr ---------- (j)

So, the final flow rates for HP & LP Separators are as follows:

	HP		LP	
	Kg/hr	m3/hr	Kg/hr	m3/hr
Oil	505762.44	659.75	381236.65	496.84
Gas	622432.97	44618.91	30345.74	4243.56
Water	827296.18	828.95	94180.04	94.368

12. RESULTS

The final mole fractions obtained for our system are as follows:

For HP Separator:
Temperature, $T=40^oC$
Pressure, $P= 20$ bar

Components	Given Input	Cal Input Mol-Fraction	Final Calculated Mole-fractions		
		z_i	X_i^L	Y_i	X_i^W
Nitrogen	0,005	0,00286402	0,000388138	0,00868272	3,24E-08
Carbon doixide	0,015	0,00859205	0,00726408	0,0210489	3,27E-05
Methane	0,5	0,286402	0,0893908	0,827152	2,32E-05
Ethane	0,07	0,0400962	0,0425698	0,0913826	2,58E-06
Propane	0,037	0,0211937	0,0433382	0,031374	1,15E-07
N-Butane	0,018	0,0103105	0,0305167	0,00759864	3,06E-09
1-Pentane	0,022	0,0126017	0,0439926	0,00384806	9,19E-11
1-Hexene	0,033	0,0189025	0,0698018	0,00267421	1,69E-11
N-Hexane	0,025	0,0143201	0,0532555	0,00172092	1,46E-12
1-Heptane	0,02	0,0114561	0,043662	0,000517412	1,46E-14
Tridecane	0,105	0,0601443	0,232557	9,98E-06	8,31E-29
Octadecane	0,082	0,0469699	0,181625	1,32E-07	1,04E-42
N-Eicosane	0,068	0,0365449	0,141313	1,43E-08	1,21E-48
Water (75 %)	0,75	0,429602	0,020325	0,00399029	0,999941
TOTAL		1,000000	1,000000	1,000000	1,000000

For LP Separator:
Temperature, $T=45^oC$
Pressure, $P= 10$ bar

Components	Given Input	Cal Input Molfraction	Final Calculated Mole-fractions		
		z_i	X_i^L	Y_i	X_i^W
Nitrogen	0,000388138	0,00031512	9,15E-05	4,10E-03	9,23E-09
Carbon doixide	0,00726408	0,00589757	0,00534419	0,0310536	2,61E-05
Methane	0,0893908	0,07257470	0,0390223	0,718881	1,20E-05
Ethane	0,0425698	0,03456160	0,0343052	0,144951	2,60E-06
Propane	0,0433382	0,03518550	0,0419778	0,059013	1,53E-07
N-Butane	0,0305167	0,02477590	0,0316661	0,0150946	4,85E-09
1-Pentane	0,0439926	0,03571680	0,0467676	0,00772344	1,68E-10
1-Hexene	0,0698018	0,05667080	0,0747542	0,00535586	3,33E-11
N-Hexane	0,0532555	0,04323710	0,0570862	0,00342993	3,06E-12
1-Heptane	0,043662	0,03544840	0,0469454	0,00101968	3,50E-14
Tridecane	0,232557	0,18880900	0,250476	1,78E-05	5,48E-28
Octadecane	0,181625	0,14745800	0,195621	2,14E-07	1,73E-41
N-Eicosane	0,141313	0,11472900	0,152203	2,25E-08	2,96E-47
Water (25 %)	0,025325	0,20462000	0,0237395	0,00935983	0,999959
TOTAL		0,999999	1,000000	1,000001	1,000000

The final calculation table showing results is as follows:

Table – 1 Separation and properties of Oil, Gas & Water

PARAMETER	HP SEPARATOR	LP SEPARATOR	UNITS
Density of gas	13.94	7.15	kg/m³
Density of oil	766.58	767.31	kg/m³
Viscosity of gas	0.01191	0.01193	CP
Viscosity of oil	1.5772	1.3784	CP
Thermal conductivity	0.10013	0.09979	W/m°K
Flow rate of gas	44618.9	4243.5	m³/hr
Flow rate of oil	659.75	496.84	m³/hr
Flow rate of water	828.95	94.368	m³/hr
API°	53.0843	52.909	

13. CONCLUSION

Numerous algorithms are adapted for designing of thermodynamic systems. Here Peng-Robinson Equation-of-State (PR) is used for Crude phase systems which is widely used in Oil & Gas industry. The calculated results using scientific EXCEL for three phase oil & gas separation are found accurate with Pro II process modelling Software.

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BIOGRAPHIES

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