

Development of Model and Simulation of a Two-Phase, Gas-Liquid Horizontal Separator

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Abstract

In recent years, the performance requirements for petroleum process plants have become increasingly difficult to satisfy. In order to understand, design and operate the complex systems in the petroleum industries at relatively low cost and with minimum risk, mathematical modelling becomes very useful. Thus, this paper proposes some developed mathematical models for a two-phase gas-liquid horizontal separator, which is valid with an accuracy of about 1.641% based on the liquid temperature values. Simulated temperature values were 300.24 K, 299.69 K and 299.14 K with the corresponding industrial values as 300.22 K, 299.67 K and 299.11 K respectively. Within the boundaries of the limitations stated, the model could be used to predict the operation of the separator at different operating conditions, to optimize the separator products and as a tool for further expansion amongst other uses.

Keywords

Gas-liquid, two phase horizontal separator, process control, model, simulation

Introduction

Crude petroleum oil as found in nature, require pre-treatment to improve the product quality and meet current quantity and specification demanded for transportation fuels and related products. These pre-treatment processes include degassing, dewatering and desalting. Incidentally, these pre-treatment processes are also complex and require sophisticated equipment and instrumentation. As a result, the petroleum producing industries are capital intensive (Meyer, 1992). In order to understand, design and operate these complex systems at relatively low cost and with minimum risk, mathematical modelling becomes very useful. Again, the cost and risk involved in testing real life engineering systems are usually prohibitive. Thus, the preferred alternative is to develop mathematical model of the system and use same for investigating its performance.

Modelling of a chemical engineering system such as a crude petroleum separating unity involves the use of fundamental physical and chemical laws (such as the laws of conservation of mass, energy and momentum) in their general form with time derivatives included to formulate the basic equations. Assumptions, which impose limitations on the model, are also made so that the model would not be too complex and impractical. The solution of the model equations may be carried out analytically or by using computerized numerical programmes. The latter is referred to as computer simulation (Imeh, 2003; Sinnott, 2001; Stephanopoulos, 1995; Luyben, 1990).

This paper describes the development of mathematical models for gas-liquid separation process and to apply the model in simulating the operation of a gas-liquid separation unit (a horizontal two-phase separator in the continuous mode) for the separation of gas from liquid petroleum.

The following limitations were applied to the developed model:

[1] It applies to a two-phase, gas-liquid separator, [2] strictly for a crude oil/natural gas system, [3] it is based on the classification of crude oil into eleven major components, [4] its tests of accuracy is done based on the temperature of the liquid (T_L) values, [5] it was generated for a system operating under normal conditions.

Description of the Gas-Liquid Horizontal Separator

A schematic diagram of the two-phase gas-liquid separator is presented in Fig 1.

The separation of a combined gas-liquid stream into separate phases can be described by the following process; the combined liquid-gas stream enters the primary separation section A (Fig. 1) which contains the inlet nozzle (which is usually tangential) or a diverter baffle. This section reduces the velocity of the fluid stream and takes advantage of the initial effects of a centrifugal or an abrupt change in direction to separate the major portion of the liquid from the combined stream. The gas then moves to the secondary section B at a lower velocity. In the secondary section B, the force of gravity is used to separate entrained droplets of liquid in the gas stream. The gas moves at relatively low velocity with little turbulence. Then the gas stream moves to the coalescing section C in which the very small droplets of liquid in the gas stream are removed by impingement on a surface where they coalesce. The section utilizes a coalescer or mist extractor, which can consist of as a series of vanes, a knitted wire mesh pad or cyclonic passage. The gas stream leaves the separator with less than 0.1 gallon per million standard cubic feet, while the liquid leaves the liquid collection section D which acts as receiver for all liquid removed from the gas in the primary, secondary and coalescing sections (Richardson and Coulson, 2001; McCabe et-al, 1993).

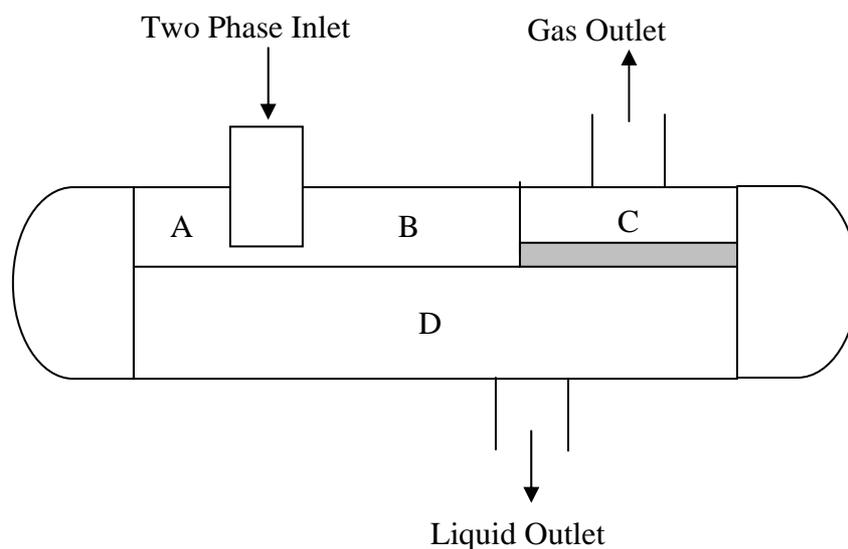


Figure 1. A two Phase gas-liquid horizontal separator

The mechanism of operation of the two-phase horizontal separator is the flash separation, where crude oil enters the separator and additional light hydrocarbon vapour are

flushed. Water is separated from the crude/water mixture, the difference in densities and the tangential configuration of the stream path acts to separate the light hydrocarbon vapour and water vapour from the crude oil. Thus chemical separation (flashing) and physical separation (gravity settling, centrifugal) methods occurs simultaneously in the separator (Imeh, 2003; Perry and Green, 1997; Smith et-al, 1996).

Development of Models for the Separator

In the development of the models for the two phase gas-liquid separator, the following assumptions were made: [1] The liquid particles are spherical in size, [2] the dynamics of the vapour phase is ignored because vapour is assumed to always be in equilibrium with the liquid, [3] the mode of settlement of the liquid particles from the gas phase is free settling i.e. the settling of individual particles is not hindered by others, [4] the pressure of the liquid and vapour in the separator are not equal, [5] the liquid density, ρ , remain constant, i.e the temperature variations are small enough that the temperature dependence of ρ can be neglected, [6] heat losses are negligible and no temperature gradients exist within the separator. Therefore, the exit temperature is equal to the temperature of the liquid in the tank, [7] variables within the separator do not vary with the length of the separator.

Applying the basic principle of conservation for the separator system under study where the total mass and energy cannot be generated neither do they disappear, so that the mass balance gives (Smith et-al, 1996; Luyben, 1990):

$$dM/dt = \Sigma \rho_i F_i - \Sigma \rho_j F_j \quad (1)$$

where ρ_i , ρ_j are the densities of inlet and outlet streams; F_i, F_j are the volumetric flow rates of inlet and outlet streams. In terms of the total mass flow rate equation (1) becomes:

$$dM/dt = W_i - W_j \quad (2)$$

$$W_i = W_o \quad (3)$$

$$W_j = W_v + W_L \quad (4)$$

where W_i , W_j are the total rate of inlet and outlet streams; W_v , W_L are the mass flow rates of vapour and liquid outlet streams; W_o is the mass flow rate of mixture in inlet stream. By assumption, neglecting the dynamics of the vapour phase we have:

$$d(\rho_L A h t) / dt = W_o - W_v - W_L \quad (5)$$

where ρ_L is the density of liquid stream; ht the height of liquid in separator; A is the area of separator. Since ρ_L and A do not varies with time, we obtain:

$$\rho_L A d(ht)/dt = W_o - W_v - W_L \quad (6)$$

$$d(ht)/dt = 1/\rho_L A [W_o - W_v - W_L] \quad (7)$$

Equation (7) gives the first required model. Now the total energy balance is given by (Imeh, 2003; Luyben, 1990):

$$dE/dt = d(U+K+P)/dt = \sum \rho_i F_i h_i - \sum \rho_j F_j h_j \quad (8)$$

where E is the total energy content; U is the internal energy; K is the kinetic energy; P is the potential energy; h_i , h_j are the enthalpies of inlet and outlet streams. Since for liquid system the change in kinetic and potential energies are negligible, that is:

$$dK/dt = dP/dt = 0 \quad (9)$$

$$dU/dt \approx dH/dt \quad (10)$$

where H is the total enthalpy of the liquid. Neglecting the vapour phase energy dynamics:

$$dH/dt = W_o h_o - W_v h_v - W_L h \quad (11)$$

where h , h_v , h_o are the specific enthalpies of the liquid, vapour and mixture streams respectively. Recall that (Smith et-al, 1996; Luyben, 1990):

$$H = \rho_L A h t C_{pL} (T_L - T_{ref}) \quad (12)$$

$$h_o = C_{p_o} (T_o - T_{ref}) \quad (13)$$

$$h_v = C_{p_v} (T_v - T_{ref}) \quad (14)$$

$$h = C_{pL} (T_L - T_{ref}) \quad (15)$$

where C_{p_o}, C_{p_v}, C_{pL} are the specific heat capacities of mixture, vapour and liquid; T_o, T_v, T_L are the operating temperatures of then mixture, vapour and liquid; T_{ref} is the reference temperature where the specific enthalpy is assumed to be zero. Substituting equations (12) to (15) into (11):

$$d(\rho_L A h t C_{pL} T_L)/dt = W_o C_{p_o} T_o - W_v C_{p_v} T_v - W_L C_{pL} T_L \quad (16)$$

$$\rho_L A C_{pL} d(ht T_L)/dt = W_o C_{p_o} T_o - W_v C_{p_v} T_v - W_L C_{pL} T_L \quad (17)$$

$$d(ht T_L)/dt = 1/\rho_L A C_{pL} \{W_o C_{p_o} T_o - W_v C_{p_v} T_v - W_L C_{pL} T_L\} \quad (18)$$

Applying the product rule in differentiation (Stroud, 1995a,b; Holman, 1989):

$$d(ht T_L)/dt = T_L d(ht)/dt + ht d(T_L)/dt \quad (19)$$

Substituting equation (7) into (19):

$$\frac{d(ht T_L)}{dt} = \frac{T_L}{\rho_L A} (W_o - W_v - W_L) + ht \frac{dT_L}{dt} \quad (20)$$

Equating equations (18) and (20):

$$\frac{1}{\rho_L C_{pL} A} (W_o C_{p_o} T_o - W_v C_{p_v} T_v - W_L C_{pL} T_L) = \frac{T_L}{\rho_L A} (W_o - W_v - W_L) + ht \frac{dT_L}{dt} \quad (21)$$

$$\frac{dT_L}{dt} = \frac{1}{ht\rho_L C_{pL} A} (W_o C_{p_o} T_o - W_v C_{p_v} T_v - W_L C_{pL} T_L) - \frac{T_L}{ht\rho_L A} (W_o - W_v - W_L) \quad (22)$$

Equation (22) gives the second required model. Considering the total mass balance (Imeh, 2003; Luyben, 1996):

$$dM_i / dt = (W_o x_{oi} / M_o^{av}) - (W_v y_i / M_v^{av}) - (W_L x_i / M_L^{av}) \quad (23)$$

$$\frac{dM_i}{dt} = \frac{d(\rho_L A h x_i) / M_L^{av}}{dt} = \frac{W_o x_i}{M_o^{av}} - \frac{W_v y_i}{M_v^{av}} - \frac{W_L x_i}{M_L^{av}} \quad (24)$$

where x_{oi} , x_i , y_i are the mole fractions of the component i in the mixture, liquid and vapour streams respectively. M_o^{av} , M_L^{av} , M_v^{av} are the average molar masses of component i in the mixture, liquid and vapour streams. Since ρ_L , A are constants equation (24) becomes:

$$\frac{\rho_L A d(hx_i) / M_L^{av}}{dt} = \frac{W_o x_i}{M_o^{av}} - \frac{W_v y_i}{M_v^{av}} - \frac{W_L x_i}{M_L^{av}} \quad (25)$$

$$\frac{d(hx_i) / M_L^{av}}{dt} = \frac{1}{\rho_L A} \left\{ \frac{W_o x_i}{M_o^{av}} - \frac{W_v y_i}{M_v^{av}} - \frac{W_L x_i}{M_L^{av}} \right\} \quad (26)$$

Recall from differentiation of product rules (Stroud, 1995; Holman, 1989):

$$\frac{d(hx_i) / M_L^{av}}{dt} = \frac{x_i}{M_L^{av}} \frac{d(ht)}{dt} + \frac{ht}{M_L^{av}} \frac{d(x_i)}{dt} \quad (27)$$

Substituting equation (7) into (27):

$$\frac{d(hx_i) / M_L^{av}}{dt} = \frac{x_i}{\rho_L A M_L^{av}} (W_o - W_v - W_L) + \frac{ht}{M_L^{av}} \frac{d(x_i)}{dt} \quad (28)$$

Equating equations (26) and (28) to obtain:

$$\frac{1}{\rho_L A} \left(\frac{W_o x_{oi}}{M_o^{av}} - \frac{W_v y_i}{M_v^{av}} - \frac{W_L x_i}{M_L^{av}} \right) = \frac{x_i}{\rho_L A M_L^{av}} (W_o - W_v - W_L) + ht \frac{d(x_i) / M_L^{av}}{dt} \quad (29)$$

$$\frac{d(x_i) / M_L^{av}}{dt} = \frac{1}{ht\rho_L A} \left(\frac{W_o x_{oi}}{M_o^{av}} - \frac{W_v y_i}{M_v^{av}} - \frac{W_L x_i}{M_L^{av}} \right) - \frac{x_i}{ht\rho_L A M_L^{av}} (W_o - W_v - W_L) \quad (30)$$

Equations (7), (22) and (30) are the developed models for the separator. In order to ensure that the degree of freedom is zero some algebraic equations were also applied (Smith et-al, 1996; Ludwig, 1994):

$$M_v^{av} = \sum y_i M_{oi} \quad (31)$$

$$y_i = \{x_{oi} k_i\} / \{1 + v(k_i - 1)\} \quad (32)$$

$$\sum y_i = 1 \quad (33)$$

$$x_i = y_i / k_i \quad (34)$$

$$M_o^{av} = \sum x_i M_{oi} \quad (35)$$

$$k_i = P_i^{sat} / P \quad (36)$$

$$W_L = W_{LZ} + Kc_1(ht_s - ht) \quad (37)$$

$$W_v = W_{vZ} + Kc_2(P_{vs} - P_v) \quad (38)$$

Equations (37) to (38) are the generated control equations, where W_L , W_v are the outlet flow rate of liquid and vapour streams; W_{LZ} , W_{vZ} are the flow rates of liquid and vapour when error is zero, ht_s , ht are the set point and measured heights of liquid. Kc_1 , Kc_2 are the proportional gain of the controller. P_{vs} , P_v are the set point and measured pressure in the separator all respectively.

Results and Discussion

The industrial and simulated results are presented in Table 1-4 and the comparison of the simulated and industrial results are presented in Table 5.

Table 1. Industrial results for the separator

W_o (kgs ⁻¹)	T_o (K)	ρ_L (kgm ⁻³)	P_v (kNm ⁻²)	T_v (K)	T_L (K)
83.567	299.67	854.61	790.834	309.67	304.11
83.500	299.67	854.61	790.834	309.67	304.11
84.835	298.56	852.05	777.044	309.67	303.00
83.301	298.56	852.05	721.886	309.67	304.11
83.591	299.11	852.05	777.044	310.22	304.11
83.801	299.11	849.49	721.886	309.67	305.22
83.678	299.67	849.49	721.886	309.67	304.11
83.695	299.67	854.61	721.886	310.22	304.11
83.647	300.22	852.05	721.886	309.67	304.11
83.337	299.11	854.61	721.886	308.56	304.11
85.670	299.67	849.49	777.044	309.67	305.22
85.572	299.67	844.36	770.149	309.67	304.67
85.766	299.67	849.49	777.044	309.67	304.11
85.812	299.67	852.05	756.360	309.67	304.67
83.889	299.67	854.61	790.834	310.22	305.22
85.510	300.22	849.49	777.044	309.11	304.11

Table 2. Simulated results for the separator

V (moles)	T _L (K)	ht _L (m)	x ₁ /M _L ^{av} · 10 ⁻⁶ (kg ⁻¹)
0.013	299.6889	1.4974	2.54
0.013	299.6889	1.4999	2.54
0.013	298.5894	1.5942	2.53
0.014	298.5835	1.4782	2.48
0.013	299.1390	1.5002	2.53
0.014	299.1390	1.5155	2.48
0.014	299.6885	1.5059	2.48
0.014	299.6882	1.5076	2.48
0.014	300.2394	1.5036	2.48
0.014	299.1362	1.4808	2.49
0.013	299.6903	1.6581	2.53
0.013	299.6896	1.6523	2.53
0.013	299.6905	1.6654	2.53
0.014	299.6910	1.6680	2.50
0.013	299.6882	1.5219	2.53
0.013	300.2407	1.6459	2.53

Table 3. Simulated results for the separator

x ₂ /M _L ^{av} · 10 ⁻⁵ (kg ⁻¹)	x ₃ /M _L ^{av} · 10 ⁻⁴ (kg ⁻¹)	x ₄ /M _L ^{av} · 10 ⁻⁴ (kg ⁻¹)	x ₅ /M _L ^{av} · 10 ⁻⁵ (kg ⁻¹)	x ₆ /M _L ^{av} · 10 ⁻⁴ (kg ⁻¹)
9.36	1.03	2.17	9.28	2.25
9.36	1.03	2.17	9.28	2.25
9.27	1.02	2.16	9.27	2.25
8.52	1.01	2.16	9.25	2.25
9.21	1.02	2.17	9.27	2.25
8.52	1.01	2.16	9.25	2.25
8.52	1.01	2.16	9.25	2.24
8.46	1.01	2.16	9.24	2.24
8.52	1.01	2.16	9.25	2.24
8.64	1.02	2.16	9.25	2.25
8.52	1.02	2.17	9.27	2.25
9.22	1.02	2.17	9.27	2.25
9.26	1.02	2.17	9.27	2.25
8.74	1.02	2.16	9.25	2.25
9.30	1.02	2.17	9.27	2.25
9.32	1.02	2.17	9.28	2.25

Eleven different components were simulated for, using the models developed. The mole fractions of these components are; x₁ is carbon dioxide; x₂ is methane; x₃ is ethane; x₄ is propane; x₅ is isobutene; x₆ is normal butane; x₇ is isopentane; x₈ is normal pentane; x₉ is normal hexane; x₁₀ is the heavy components of crude oil and x₁₁ is water.

Table 4: Simulated results for the separator

$x_7/M_L^{av} \cdot 10^{-4}$ (kg ⁻¹)	$x_8/M_L^{av} \cdot 10^{-4}$ (kg ⁻¹)	$x_9/M_L^{av} \cdot 10^{-4}$ (kg ⁻¹)	$x_{10}/M_L^{av} \cdot 10^{-3}$ (kg ⁻¹)	$x_{11}/M_L^{av} \cdot 10^{-3}$ (kg ⁻¹)
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.65	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04
1.64	1.74	2.15	3.83	5.04

From the tables of results (Tables 1-4) presented, the variables solved for are fairly constant despite the varying values of disturbances like W_o , T_o etc. The moles of the chemical species (V) had values over a range of 0.013 to 0.015, the temperature of liquid T_L , has 298.59 to 300.24 K, while the height of liquid in the separator, ht_L , has 1.4974 to 1.6680 m. This infer that the control laws given by the control equations were effective in keeping output variable stable (Imeh, 2003; Stephanopoulous, 1995). It is also remarkable that the values of the variables obtained are realistic and logical. Values of V obtained in the range of 0.013 to 0.015 shows that the vapour phase of the separator's discharge is minute.

This corroborates literature findings (Odigure et-al, 2003; Odigure, 1998; Erika et-al, 1988) and further confirmed by industrial data. The values of T_L obtained are equally realistic values for crude oil under the given conditions.

A good relationship was observed between the disturbance variables and the output variables. W_o is seen to be directly proportional to the height, ht (Tables 1 and 2). The highest value of W_o of 85.812 kgs⁻¹ has a corresponding highest value of ht , as 1.6680 m and vice versa for the lowest values. T_o was seen to be directly proportional to T_L where a high value of 300.22 K corresponds to an equally high value of T_L of 300.24 K and vice versa. In the same vein, V was observed to vary directly with P_v and indirectly with T_v (Luyben, 1990; Seborg, 1989). Table 5 present the deviation calculated between the industrial and simulated, this deviation obtained was 4.996 giving a mean error value of 1.641 % based on T_L values.

Table 5: Comparison of Simulated and Industrial temperatures

Simulated (K)	Industrial (K)	Absolute Deviation (K)	Error (%)
299.69	304.11	4.42	1.453
299.69	204.11	4.42	1.453
299.56	303.00	3.41	1.125
298.56	304.11	5.52	1.815
299.14	304.11	4.97	1.634
299.14	305.22	6.08	1.992
299.69	304.11	4.42	1.453
299.69	304.11	4.42	1.453
300.24	304.11	3.87	1.273
299.14	304.11	4.97	1.634
299.69	305.67	5.53	1.812
299.69	304.69	4.42	1.435
299.69	304.67	4.98	1.635
299.69	304.67	4.98	1.635
299.69	305.22	5.53	1.812
300.24	304.11	3.87	1.273

On one hand, the model proved valid because of the low percentage error generated while on the other hand, the error generated could be traced to some of the assumption made. Parameters such as the controller gain, area of separator, and steady state values could have contributed. The numerical integration method used in solving differential equations could also have introduced truncation errors in the process of iteration (Stroud, 1995).

Conclusion

Modelling is a very important tool used in the oil industries to describe operating systems. This paper proposes some developed mathematical models for a two-phase gas-liquid horizontal separator, which is valid with an accuracy of about 1.641% based on the liquid temperature values. Within the boundaries of the limitations stated, the model could be used to predict the operation of the separator at different operating conditions, to optimize the separator products and as a tool for further expansion amongst other uses.

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