Recovery of flared gas through crude oil stabilization by a multi-staged separation with intermediate feeds: A case study

Djebri Mourad*, Otmanine Ghazi**,†, and Bentahar Noureddine**

*Sonatrach Direction Production, Région de TFT, BP66, IN AMENAS, Algeria
**Faculty of Hydrocarbons & Chemistry, University M'Hamed Bougara,
Avenue de l'Indépendance, 35000 - Boumerdès, Algeria

Abstract−Since the birth of the oil industry, flaring has been used upstream to depressurize eruptive wells and downstream to burn excess gases in refining and petrochemical plants and also in associated and natural gas treatment plants. Unfortunately, flaring produces a great number of harmful by-products such as dangerous particles, volatile organic compounds, polycyclic aromatic and many other compounds just as harmful. The separation of gas and oil phases remains the most important stage in the so-called surface production. Given the high gas oil ratio (GOR) of Algerian crude oil, the separation of this gas is carried out in three or four stages. However, the choice of the optimal number of stages of separation and intermediate values of pressure requires a rigorous computation wherein the use of a simulator is more than necessary to make possible the optimization of the separation process. The present work was performed on a new separation and compression unit in an area where all the associated gas production is being flared despite the new environmental laws. Our approach consists of first simulating the separation process with the most appropriate thermo dynamical model. The intermediate separation pressure values can be determined by empirical correlations such as the method of equal pressure ratio. In our computations we have opted for a graphical method, specifically the method of minimum compression energy, that requires rigorous calculations entailing therefore the use of the Hysis simulator. This treated gas may be valorized as a raw material for the petrochemical industry or compressed and reinjected into the reservoir in order to maintain the rate of oil production. It remains that one important way of valorizing this associated gas is to transform it into liquid through a process known as gas to liquid (GTL) technology.

Key words: Separation, Oil, Associated Gas, Compression, Flare

INTRODUCTION

In the oil and gas industry flaring is seen to be a safe and reliable method for disposal of residual gas to be burned in the open air using a special device called a torch. Gas flaring is needed to depressurize eruptive wells and hydrocarbon plants through burning excess gases. But this measure that enhances plant safety is not without danger to the environment.

Indeed, the burned gases in addition to CO_2 and water vapor, will produce dangerous compounds such as particles, volatile organic compounds (VOCs) such as benzene, toluene and xylene; VOCs have been lately the object of attention of a good number of research workers [1,2]. Other dangerous compounds also emitted are the polycyclic aromatic hydrocarbons (PAHs), sulfur compounds in small amounts such as carbon disulfide (CS_2) and carbonyl sulfide (COS).

A number of international standards have been established in conjunction with the installation and operation of torches. However, these standards have not been abided by.

During the sixties the gases produced by various wells and production plants had no market; therefore they were inevitably sent to the torch.

According to the World Bank [3], the annual volume of natural gas flared or vented in the world for the year 2003 amounted to more

than 100 billion cubic meters, exactly (108 Gcm) (Fig. 1), which represents the annual gas consumption of France and Germany combined. This amount of gas flaring level has remained constant over the past 20 years. Flaring in Africa alone accounts for almost 35% of global flaring (38 Gcm) [4] as shown in Fig. 1. In Algeria alone flaring amounts to (5 Gcm) [5].

Gas flaring has become a public concern and a priority issue for public authorities because it is first a waste of a non-renewable source in addition to environmental problems due to gas emissions that
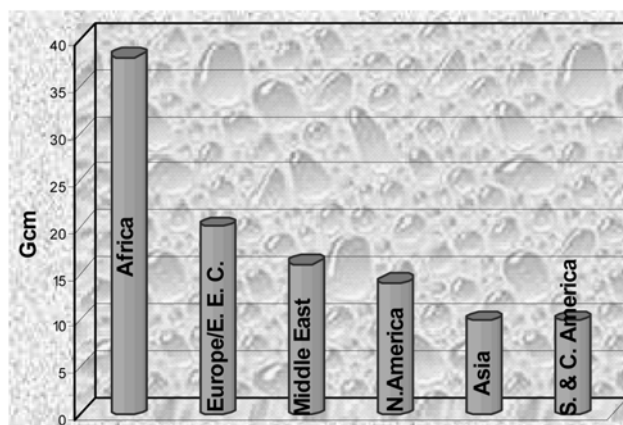

Fig. 1. Gas flared throughout the world in Gcm.

†To whom correspondence should be addressed.
E-mail: oth_gha@umbb.dz

produce greenhouse effects; it also represents definite risks to human health.

According to the Office for Information on Climate Change (BICC) of the program of the United Nations Environment (UNEP) [6], it is estimated that global emissions of carbon dioxide due to flaring reached 202 million tons in 1989, approximately 0.8% of anthropogenic emissions origins of carbon dioxide.

Indeed, we are witnessing an appreciable increase in a certain types of diseases such as asthma, cardiovascular diseases and various cancers. Scientists predict that rising levels of greenhouse gases will result in an important increase of global warming. This phenomenon could cause disastrous changes on the environment, such as severe storms, desert advance and melting ice caps that will raise sea levels and therefore submerge coastal regions.

Nonetheless, world organizations are striving to implement a global policy that seeks to reduce gas emissions due to flaring.

It was not until 1997 and after many preliminary meetings and conferences that representatives of governments worldwide have attended and responded to the increasing pressure from the international community to sign the Kyoto Protocol [7] setting as a target a reduction of 10 billion cubic meters of gas flared for the period between 2008 and 2012. It is a clear sign that the international community has the will to face reality and start taking concrete measures to minimize the risks that may take place on climate change, health and the environment especially after the signing of Russia despite the refusal of the USA who are a major source of greenhouse effects gases.

It is high time to think seriously about alternative solutions to gas flaring that are less wasteful of energy and less harmful to the environment. The solutions are many and among them are:
- Re-injection of these gases in the oil fields to enhance recovery of crude oil if the characteristics of the reservoir permit;
- Re-injection of these gases in wet gas fields to maximize the recovery of liquids;
- Oil production by gas lift system;
- Use of associated gases as fuel gas;
- Construction of a transport system for collecting gases to be shipped to a treatment plant before their subsequent use;
- The technology of converting gas to liquid "GTL" (gas to liquid) conversion, which is beginning to be applied on a large scale.

Finally, another solution for eliminating residual gases would be to burn them in incinerators that are preferable to torches because less pollution is caused and the heat can be recovered for some useful purpose (co-generation).

The following work is focused on a study of a classic solution, which is the most widespread, that consists of collecting and compressing associated gases in order to send them to a plant where they are treated and thus become ready for being marketed.

THEORETICAL BACKGROUND

The fluid in the reservoir is a mixture of gaseous liquid, hydrocarbons and water. Initially, the mixture is in a state of equilibrium which depends on the composition as well as the existing conditions of pressure and temperature in the reservoir. During exploitation the balance is broken by the pressure losses in the reservoir rock on the one hand and in the collection pipes on the other. The effluent undergoes successive decompressions until its injection into the treatment plant, the decompressions taking place with a drop in temperature.

This pressure drop will release a more or less important quantity of gas. If no product is withdrawn on the course, the product total mass will remain constant. Only the ratio in volume of the gaseous phase over the liquid phase varies; this is said to be a flash. In this case, the equilibrium between the gaseous phase and the liquid phase is reached. Gas will drive a non negligible amount of heavy fractions, so it must then be stabilized by reducing the pressure. This operation known as flash separation will allow the production of a maximum volume of gas on the one hand and less liquid (condensate or oil) on the other hand [8].

If the gas mixture (consisting of minor constituents) is racked continuously whenever it is formed, there is talk of separation differential [8]. In this operation, a maximum of intermediate and heavy components remain in the liquid phase because the balance is not achieved, so light components lead the heaviest in the vapor, resulting in a maximum volume of liquid.

Furthermore, if we had the opportunity to move the effluent through an infinite number of dividers placed in series (multi-staged separation) and operating at uniformly decreasing pressure between the bubble pressure and the storage pressure and if each separator was unloading gas released at its own pressure, this would then be the ideal differential vaporization wherein the state of phase equilibrium for each stage would be perfectly ensured.

The separation which tends to move closer to the ideal conditions of differential vaporization provides less gas and more liquid.

The recovery of the liquid phase increases with the number of stages of separation. In practice, this number is limited between 2 and 4 [9] and depends on the gas oil ratio (GOR) and the wellhead pressure.

Three modes of separation should be considered:
- Two stages: low GOR, low wellhead pressure.
- Three stages: average GOR, intermediate pressure.
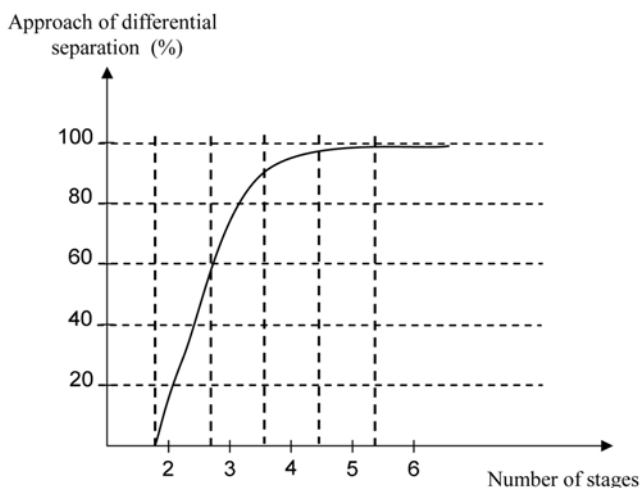- Four stages: high GOR, output pressure or pressure after high compression.

Fig. 2. The recovery rate of the liquid depending on the number of stages.

Table 1. The recovery rate of the liquid depending on the number of stages

Number of stages	Approach of differential separation
2	0
3	57
4	90
5	96
6	98.5

Table 2. Approximate guide for choosing the number of separation stages

Pressure of the first stage separation [bars]	Number of stages
3-10	1
10-22	1-2
22-35	2
35-49	2-3
>49 (flow of effluent>100,000 bpd	>3

Fig. 2 and Table 1 show the recovery rate of liquid depending on the number of stages.

The final storage tank is generally maintained at a pressure close to atmospheric pressure and is the last separation stage.

The three-stage separation most often represents the economic optimum. It helps to recover from 2 to 12% more liquid in comparison with a two stage separation and in some cases up to 25%.

However, the choice of the number of separation stages and the calculation of the optimum value of the intermediate pressure being very complicated program requires a computing program using a model based on equations of state.

A typical example with practical data is given in Table 2.

One should note that it is illustrative and can in no way be a substitute for calculating a flash [10].

The choice of pressure staging was often based on the assumption of equal pressure ratio (Campbell 1981) [11], so that the pressure ratio of R_P in two successive stages is constant:

$$R_P = \left(\frac{P_{i-1}}{P_i}\right) = \left(\frac{P_i}{P_n}\right)^{1/n} \qquad (1)$$

Where P_i: pressure of stage (i); P_{i-1}: pressure of stage (i−1); P_n: pressure of stage (n) and n: number of stages.

The pressure in the first stage (high pressure: HP) and the pressure on the last stage (usually near the atmospheric pressure) being known, it only remains to deduce the pressure of each stage.
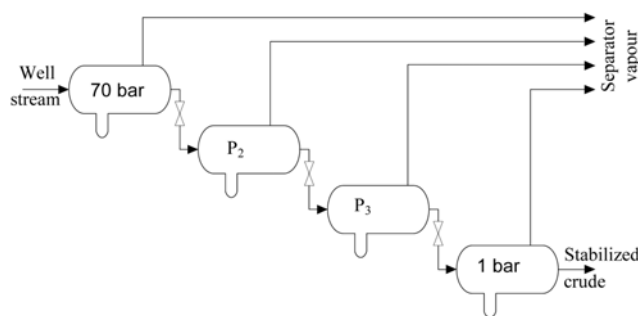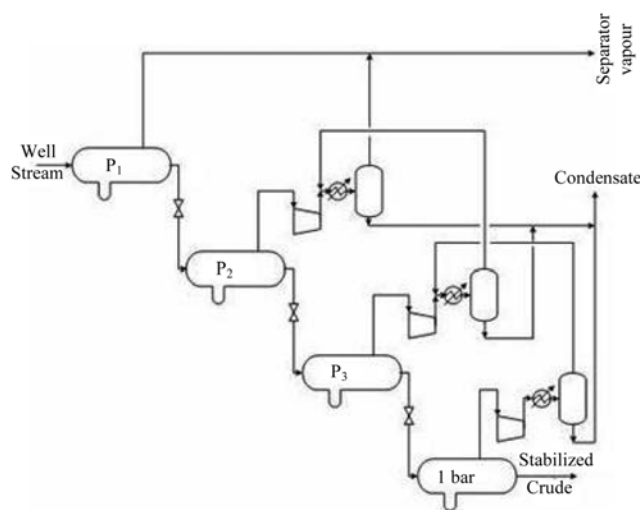
Whinery and Campbell (1958) [8] have studied the three stage separation for different types of effluents and have established an expression relating the pressure of the first stage (P_1) to that of the second stage (P_2), the pressure of the third stage (P_3) being set at atmospheric pressure value.

They have found the following:

For fluids with a density greater than 1, we have:

$$P_2 = AP_1^{0.686} + \frac{(A-0.057)}{0.0233} \qquad (2)$$

For fluids with a density less than 1, we have:

Fig. 3. Multi-stage separation.

Fig. 4. Multi-stage separation with compression of produced gas.

$$P_2 = AP_1^{0.765} + \frac{(A+0.028)}{0.012} \qquad (3)$$

The value of 'A' is a function of the third stage pressure, generally the storage pressure.

Currently, with the use of simulators, the quantities of steam and liquid recovered for a fixed pressure are quickly determined by using a model based on the equation of state relevant to this type of separation. This helps optimize the number of stages and the pressure value set for each separator.

The four-stage separation scheme shown in Fig. 3 has been studied through simulation in order to maximize the recovery of oil [12].

The same study proposes, for oil recovery, replacing the last separating ball by a stabilizer column. But in the case of associated gas, this choice is considered too expensive and requires an additional investment; hence burning it is less costly.

Flaring is a waste of energy and a source of pollution to the environment; among the solutions proposed for energy conservation is the compression of the gas produced from a separation which is then sent or re-injected into the reservoir.

In this case, one must not only maximize the recovery of oil, but also minimize compression energy.

The scheme of a multi-staged separation (4 stages) with compression of gas produced is illustrated in Fig. 4 [12].

The pressure in the first stage (high pressure: HP) and the pressure in the last stage (usually near atmospheric pressure) being known,
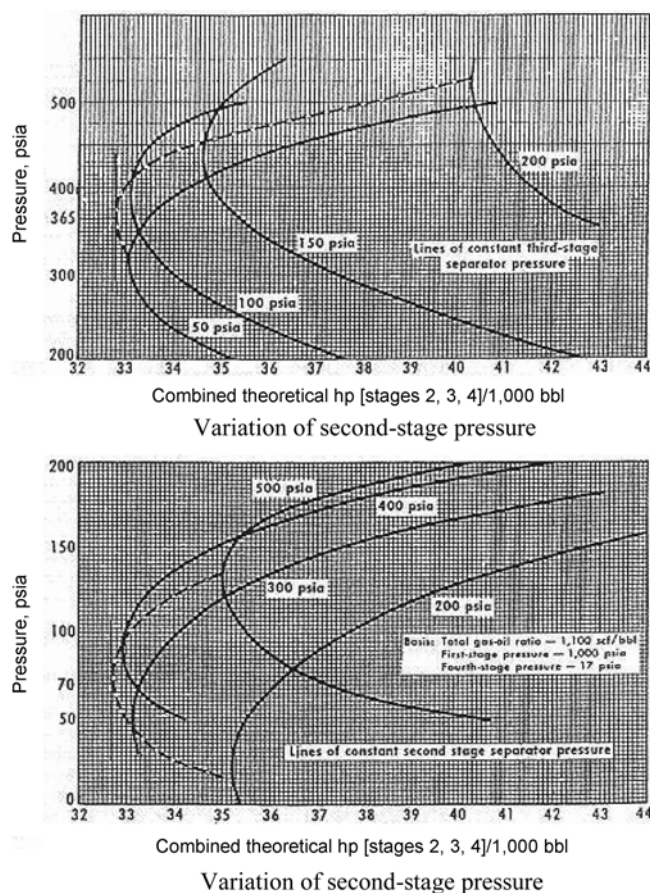
Fig. 5. Minimization of energy compression for a 4-stage separation (Kryska et al., 1976).

it only remains to deduce the pressure in the intermediate stages. To do so we must study the variation of the total compression energy as a function of pressures P_2 and P_3.

Several authors have studied techniques for minimizing compression energy by setting values for intermediate pressure separation.

Kryska and al [13] have developed a simulation of a multi-staged separation process (4 stages) with compression of a gas produced from crude with a GOR of 1,100 scf /bbl. This study has shown the effect of the second and third stage on compression energy. The results are shown in Fig. 5.

PROBLEM DESCRIPTION

Associated gases present the main source of flaring. To this end we sometimes have to separate oil and gas effluents (gas condensate or associated gas) in the same battery of separation, which makes it possible to reduce investment in surface equipment.

A real case has been considered that is the building of a new unit of separation and compression in Southern Algeria (USC), a region very rich in gas with a few oil pockets. Several wells feed the unit, hence the need to carefully determine loads and supply conditions.

Gas effluents with condensate are mixed with oil effluents for feeding the unit in order to make the recovered gas compression (rich in condensate) cost effective.

There are two production phases: phase (1) summer & winter

cases, phase (2) summer & winter cases dictated by the oilfield evolution; from where the design of flexible equipment which can be used in both phases is more than necessary.

By separating oil from associated gases, we tend to maximize oil production by a multi-staged separation where the choice of the number of stages and intermediate pressures can be made on the basis of the methods mentioned above.

The treatment of gas effluent with condensate and oil in the same treatment plant as in the case of the USC affects the method of minimizing energy compression, where we tried to interpret the influence of intermediate supplies on the shape of the energy compression curve depending on the pressure.

Crude oil is a complex mixture of gaseous and liquid hydrocarbons, water and other contaminants, which leads the separation of phases to a liquid-liquid-vapor that requires rigorous computations; hence, the need to use a simulator to help optimize the separation process.

The collection of effluents based on the localization and characteristics of wells (mainly operating conditions) has allowed obtaining four feeds at the separation unit input with the characteristics given in Tables 3A, 3B, 3C, 3D and 3E. The data given are the results of a simulation of the wells network collection system using PIPESIM dedicated software.

The modelling of a physical system is based on the knowledge of a certain number of properties of pure and binaries bodies. These properties serve as a basis for the determination of thermodynamic properties of transfer through and equilibrium between phases. Thus, the quality of results is directly linked to the data of pure and binary bodies and to the adequate models used to represent the thermodynamic behavior of the system considered.

The choice of the thermodynamic model is primarily dictated by the need for analytical rigor and the reliability and validity of the results. The thermodynamic models are numerous and their scope of application is very diverse.

For complex mixtures of oil type fraction, it is not always possible to accurately analyze their characteristics in order to deduce the nature and composition of each component. Also, in order to

Table 3A. Characteristics of pseudo components

Pseudo components	Specific Gravity 60 °F	Molecular weight	Boiling temperature (°C)
C_7	0.711	99	91
C_8	0.727	110	112
C_9	0.744	124	139
C_{10}	0.758	139	163
C_{11}	0.768	152	184
C_{12+gas}	0.807	223	279
C_{7-RC}	0.702	98	88
C_{8-RC}	0.720	112	114
C_{9-RC}	0.736	126	140
C_{10-RC}	0.750	142	166
C_{11+RC}	0.800	200	252
C_{12+oil}	0.879	295	363
C_{12+A6}	0.872	317	378
C_{12+SE8}	0.884	280	352
C_{12+SE9}	0.896	354	400

Table 3B. Characteristics of feeds Phase 1 - summer case

Feeds	HP	MP	LP	VLP
Temperature (°C)	50.0	50.0	50.0	50.0
Pressure (Bar)	91.0	42.0	20.0	6.0
Flow (K mole/h) (Kg/h)	17892.3	4476.5	2621.0	822.4
	368977.0	97946.0	56051.0	23267.0
Heat flow (MJ/h)	−2619407.3	−408062.2	−421737.3	−135323.5
Flow vapour phase (m^3/h)	2870.7	2482.9	2060.8	2264.0
Flow liquid phase (m^3/h)	47.5	15.6	11.0	10.1
Flow aqueous phase (m^3/h)	100.9	1.4	17.8	4.9
Composition	% molar			
N_2	0.0038	0.0111	0.0031	0.0032
CO_2	0.0036	0.0192	0.0030	0.0031
H_2S	0.0000	0.0000	0.0000	0.0000
Methane	0.5755	0.8050	0.5388	0.5228
Ethane	0.0536	0.0768	0.0500	0.0512
Propane	0.0196	0.0252	0.0068	0.0110
i-Butane	0.0056	0.0075	0.0021	0.0035
n-Butane	0.0070	0.0074	0.0026	0.0067
i-Pentane	0.0035	0.0044	0.0006	0.0027
n-Pentane	0.0023	0.0027	0.0005	0.0034
C_6	0.0036	0.0043	0.0010	0.0049
C_7	0.0030	0.0017	0.0011	0.0062
C_8	0.0023	0.0015	0.0012	0.0061
C_9	0.0015	0.0010	0.0011	0.0052
C_{10}	0.0011	0.0007	0.0010	0.0049
C_{11}	0.0006	0.0005	0.0009	0.0035
C_{12+gaz}	0.0004	0.0024	0.0000	0.0000
C_{7-RC}	0.0000	0.0020	0.0000	0.0000
C_{8-RC}	0.0000	0.0016	0.0000	0.0000
C_{9-RC}	0.0000	0.0011	0.0000	0.0000
C_{10-RC}	0.0000	0.0009	0.0000	0.0000
C_{11+RC}	0.0000	0.0023	0.0000	0.0000
C_{12+oil}	0.0013	0.0000	0.0091	0.0000
C_{12+A6}	0.0000	0.0000	0.0005	0.0000
C_{12+SE8}	0.0002	0.0000	0.0000	0.0219
C_{12+SE9}	0.0003	0.0000	0.0000	0.0024
Water	0.3114	0.0207	0.3766	0.3375

represent such mixtures, the notion of pseudo components is introduced. Pseudo components play, in an oil cut, a role equivalent to that of a pure body in a defined mixture. It then becomes necessary for any thermodynamic computations to know the intrinsic properties of these pseudo components. These calculations are possible through the use of specific correlations that do not require knowledge of the chemical structure of the pseudo components.

It is therefore necessary to provide two of the three following properties:
- Boiling temperature at atmospheric pressure;
- Specific gravity or API degree;
- Molecular weight.

The composition of the effluent of each well and the conditions of pressure and temperature are obtained from the results of tests carried out on wells on-site.

In our case, it is about simulating a three stage separation process under equilibrium liquid-liquid-vapor (LLV) to produce an aqueous phase (water), a liquid phase (oil) and a vapor phase.

Models like UNIQUAC, Margules, UNIFAC and NRTL are recommended for LL equilibrium calculations, whereas the equations of state like SRK, PR, and BWRS etc., are recommended for working out the LV equilibrium.

This software is different from the broad field of application, the complexity of operating, flexibility and time programming.

The simulator used for our calculations is HYSYS software [14]. A large database on pure body is incorporated into the software.

The Peng-Robinson (PR) equation [15] is used for petrochemical, petroleum and gas applications within a wide range of temperature and pressure values (T>−271 °C et P<1,035 bars). It adequately describes both single phase and multi-phase systems. In addition,

Table 3C. Characteristics of feeds Phase 1 - winter case

Feeds	HP	MP	LP	VLP
Temperature (°C)	39.5	8.7	30.0	-5.0
Pressure (Bar)	91.0	42.0	20.0	6.0
Flow (K mole/h) (Kg/h)	17863.6	4476.5	2621.0	822.4
	368977.0	97946.0	56051.0	23267.0
Heat flow (MJ/h)	−2629598.5	−419560.0	−425337.1	−139294.4
Flow vapour phase (m³/h)	2680.0	1998.2	1900.8	1778.2
Flow liquid phase (m³/h)	55.7	23.4	11.3	10.9
Flow aqueous phase (m³/h)	100.0	1.6	17.6	4.8
Composition	% molar			
N_2	0.0038	0.0111	0.0031	0.0032
CO_2	0.0036	0.0192	0.0030	0.0031
H_2S	0.0000	0.0000	0.0000	0.0000
Methane	0.5755	0.8050	0.5388	0.5228
Ethane	0.0536	0.0768	0.0500	0.0512
Propane	0.0196	0.0252	0.0068	0.0110
i-Butane	0.0056	0.0075	0.0021	0.0035
n-Butane	0.0070	0.0074	0.0026	0.0067
i-Pentane	0.0035	0.0044	0.0006	0.0027
n-Pentane	0.0023	0.0027	0.0005	0.0034
C_6	0.0036	0.0043	0.0010	0.0049
C_7	0.0030	0.0017	0.0011	0.0062
C_8	0.0023	0.0015	0.0012	0.0061
C_9	0.0015	0.0010	0.0011	0.0052
C_{10}	0.0011	0.0007	0.0010	0.0049
C_{11}	0.0006	0.0005	0.0009	0.0035
C_{12+gaz}	0.0004	0.0024	0.0000	0.0000
C_{7-RC}	0.0000	0.0020	0.0000	0.0000
C_{8-RC}	0.0000	0.0016	0.0000	0.0000
C_{9-RC}	0.0000	0.0011	0.0000	0.0000
C_{10-RC}	0.0000	0.0009	0.0000	0.0000
C_{11+RC}	0.0000	0.0023	0.0000	0.0000
C_{12+oil}	0.0013	0.0000	0.0091	0.0000
C_{12+A6}	0.0000	0.0000	0.0005	0.0000
C_{12+SE8}	0.0002	0.0000	0.0000	0.0219
C_{12+SE9}	0.0003	0.0000	0.0000	0.0024
Water	0.3114	0.0207	0.3766	0.3375

several cases and applications of three phase separations similar to our case and within the range of applications of PR equation (T & P) are done by using the quoted equation, which leads us to say that the PR equation is the most appropriate one for our case.

$$P = \frac{RT}{V-b} - \frac{a(T)}{V(V+b)+b(V-b)} \qquad (4)$$

Where:

$$a = 0,45724 \frac{R^2 T^2}{P_c} \alpha(T_R) \qquad (5)$$

$$b = 0,0778 \frac{RT_c}{P_c} \qquad (6)$$

$$\alpha^{0,5} = 1 + k(1 - T_r^{0,5}) \qquad (7)$$

$$k = 0,37646 + 1,54226\omega - 0,26992\omega^2 \qquad (8)$$

Given that the pressure of the first stage is set by the pressure of the wellhead, the pressure losses and that of the last stage, which is the storage pressure, it is therefore necessary to determine the number of stages and the intermediate pressures using the Kryska method, the method of minimal compression energy (Fig. 5).

The most important flow represented by the HP collection determines the intermediate pressure of separation and the same pressure determines the number of separation stages, according to values given in Table 1.

RESULTS AND DISCUSSION

For a simulation run we first have to define the pseudo compo-

Table 3D. Characteristics of feeds Phase 2 - summer case

Feeds	MP	LP
Temperature (°C)	50.0	50.0
Pressure (Bar)	42.0	20.0
Flow (K mole/h) (Kg/h)	17924.8	3545.2
	366738.0	70081.0
Heat flow (MJ/h)	−2546144.3	−426499.3
Flow vapor phase (m³/h)	7193.7	3648.0
Flow liquid phase (m³/h)	42.9	4.1
Flow aqueous phase (m³/h)	95.4	12.4
Composition	% molar	
N_2	0.0037	0.0080
CO_2	0.0035	0.0053
H_2S	0.0000	0.0000
Methane	0.5985	0.6871
Ethane	0.0548	0.0650
Propane	0.0161	0.0161
i-Butane	0.0047	0.0047
n-Butane	0.0060	0.0047
i-Pentane	0.0027	0.0021
n-Pentane	0.0020	0.0013
C_6	0.0029	0.0021
C_7	0.0025	0.0002
C_8	0.0021	0.0002
C_9	0.0014	0.0002
C_{10}	0.0011	0.0002
C_{11}	0.0007	0.0002
C_{12+gaz}	0.0001	0.0000
C_{7-RC}	0.0000	0.0012
C_{8-RC}	0.0000	0.0009
C_{9-RC}	0.0000	0.0006
C_{10-RC}	0.0000	0.0004
C_{11+RC}	0.0000	0.0003
C_{12+oil}	0.0005	0.0019
C_{12+A6}	0.0000	0.0000
C_{12+SE8}	0.0018	0.0000
C_{12+SE9}	0.0004	0.0000
Water	0.2947	0.1974

Table 3E. Characteristics of feeds Phase 2 - winter case

Feeds	MP	LP
Temperature (°C)	26.7	16.1
Pressure (Bar)	42.0	20.0
Flow (K mole/h) (Kg/h)	17924.8	3545.2
	366738.0	70081.0
Heat flow (MJ/h)	−2573856.5	−433685.3
Flow vapor phase (m³/h)	6420.7	3167.2
Flow liquid phase (m³/h)	53.5	5.6
Flow aqueous phase (m³/h)	94.3	12.4
Composition	% molar	
N_2	0.0037	0.0080
CO_2	0.0035	0.0053
H_2S	0.0000	0.0000
Methane	0.5985	0.6871
Ethane	0.0548	0.0650
Propane	0.0161	0.0161
i-Butane	0.0047	0.0047
n-Butane	0.0060	0.0047
i-Pentane	0.0027	0.0021
n-Pentane	0.0020	0.0013
C_6	0.0029	0.0021
C_7	0.0025	0.0002
C_8	0.0021	0.0002
C_9	0.0014	0.0002
C_{10}	0.0011	0.0002
C_{11}	0.0007	0.0002
C_{12+gaz}	0.0001	0.0000
C_{7-RC}	0.0000	0.0012
C_{8-RC}	0.0000	0.0009
C_{9-RC}	0.0000	0.0006
C_{10-RC}	0.0000	0.0004
C_{11+RC}	0.0000	0.0003
C_{12+oil}	0.0005	0.0019
C_{12+A6}	0.0000	0.0000
C_{12+SE8}	0.0018	0.0000
C_{12+SE9}	0.0004	0.0000
Water	0.2947	0.1974

nents according to their properties, namely specific gravity, molecular weight and boiling temperature, as described in Table 3A. Then the Peng-Robinson equation is used to do the computation. The binary interaction parameters are estimated by the HYSIS simulator on the basis of the PR equation while setting the non hydrocarbon - hydrocarbon coefficients to zero.

Computations have been done for two cases: the summer and the winter cases each with its specific conditions. Taking into account those specific conditions the following simulation runs have been achieved:

1. Simulation of a 4 stage separation process with intermediate feed at each stage as shown in Fig. 6.

The pressure in the first separator is determined by the incoming fluid from wells; it has a value of 91 bars. The pressure in the last separator is determined by the stabilization section; it is 6 bars. There remains to determine the pressure in the second and third separators.

2. Study of the variation of the total compression energy as a function of values of intermediate pressures using data from Phase 1 summer case. Results obtained by simulation are represented in Figs. 7(a) and (b).

One notes that the shape of the curves shown in Figs. 7(a) and (b) is not like those shown in Fig. 5. In the present case, energy keeps on decreasing with increasing pressure. On the other hand, as indicated in Fig. 5, energy decreases until it reaches a minimum value and then starts increasing (parabolic shape with a minimum).

The difference between our results and the Kryska study results can be explained by the fact that in our case we have one additional feed coming from the well for each separating ball as shown in Fig. 6, whereas Kryska used only one feed for the entire scheme, connected to the first separator. The HP collection point being at
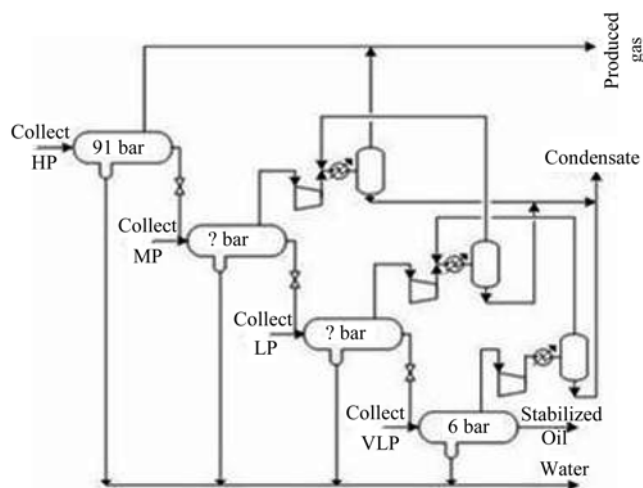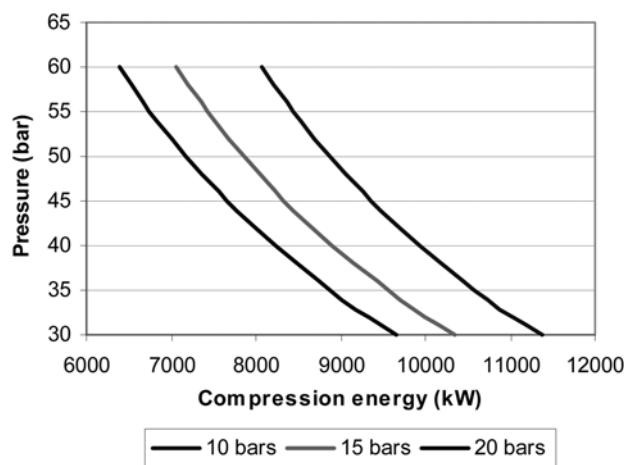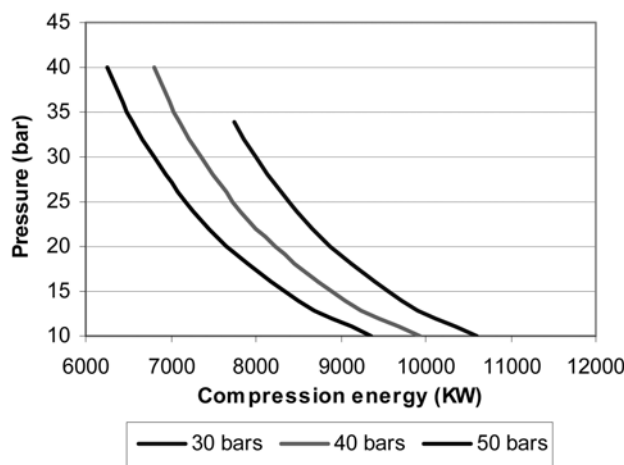
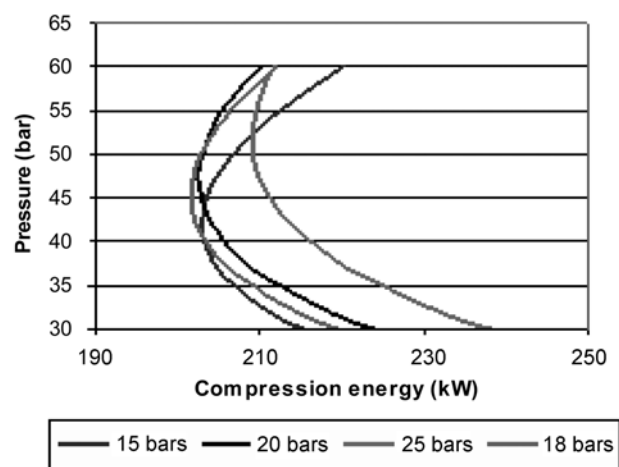Fig. 6. 4 Separation stages with intermediate feeds.
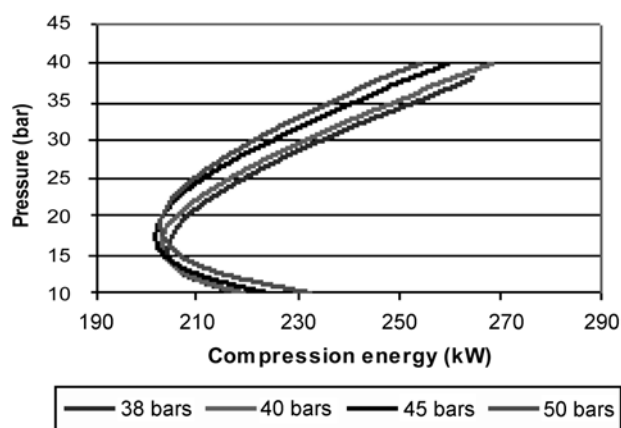

(a) Variation of second-stage pressure

(b) Variation of third-stage pressure

Fig. 7. Variation of the total compression energy as a function of intermediate pressures (all collect).

(a) Variation of second-stage pressure

(b) Variation of third-stage pressure

Fig. 8. Variation of the total compression energy as a function of intermediate pressures (only collect HP).

3. A 4 stage separation process with only one feed (HP collection) at the first stage is simulated with the variation of the total compression energy as a function of the values of intermediate pressures. Results are shown in Figs. 8(a) and (b).

Curves obtained and plotted in Fig. 8 are similar in shape with those obtained in Kryska's work, which indicates that intermediate feeds from each separator have an effect on the shape of the compression energy. The simulation process with intermediate feeds taken into account will yield a curve of compression energy as a function of pressure exhibiting an inverse proportionality relation as shown in Fig. 7. However, this is not the objective of the minimum compression energy method, which seeks to determine the value of the pressure needed to achieve minimum compression energy, i.e., a minimum point as shown in Fig. 8, and this is only possible in the case of one sole feed.

The method of minimum energy compression (MEC) yields the following values for intermediate pressures: P_2=42 bars and P_3=20 bars.

Thus, the separation process takes place in four stages with P_1= 91 bars, P_2=42 bars and P_3=20 bars and P_4=6 bars as depicted in the scheme shown in Fig. 9.

91 bars, the curve of total compression energy as a function of pressure is plotted and shown in Fig. 8.
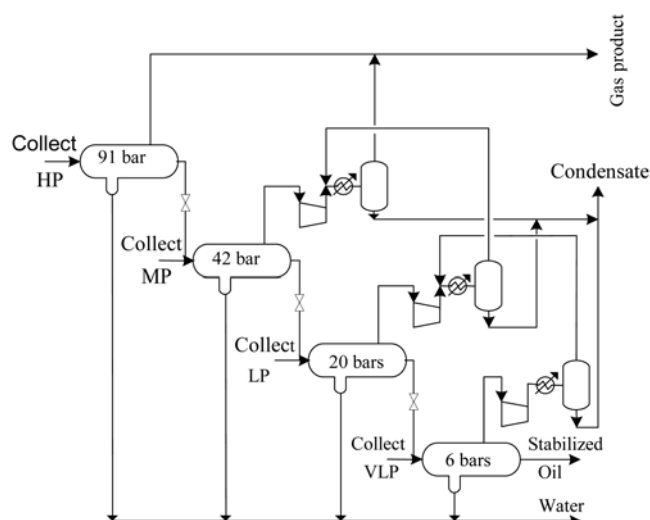
Fig. 9. Figure simplified method of separation.

AMOUNT OF RECOVERED GAS

The separation and compression unit is supplied by a mixture of gas condensate and oil. The gas produced in this center is therefore rich in condensate and LPG which can be recovered by another Process plant.

With the help of HYSYS simulator, the power flow diagram (PFD) of our proposed scheme shown in Fig. 9 is implemented by using the necessary data to carry out computations. The quantities of produced gases are estimated and shown in Table 4.

PFD simulations results in terms of produced gas and oil are shown in Table 5.

There was separation of each collection in separate batteries; it is therefore to be addressed:
- The collection HP 4 stage separation 91 bars, 42 bars, 20 bars and 6 bars.
- The collection MP in 3 stage separation 42 bars, 20 bars and 6 bars.
- The collection LP in 2 stage of separation 20 bars and 6 bars.
- The collection VLP in a single-stage separation 6 bars.

The simulation results of these separations are shown in Table 6.

Comparing the results of Tables 5 and 6 indicates an increase of 0.07% in oil production when it treats each collection separately. Of course this increase in volume of oil is the result of a decrease in volume of gas produced. This can be explained by the introduction of supplies at intermediate stages.

In making two effluents contact, the system (both effluents) tends to go to a steady state; there will be a change in concentration of constituents of the two effluents. As the two streams, one in liquid

Table 5. Simulation results with consideration of intermediate supplies

Products	Gas product	Oil product
Temperature (°C)	53.1	47.4
Pressure (Bar)	89.0	6.0
Flow (K mole/h) (Kg/h)	18604.0	320.4
	370873.8	51270.4
Heat flow (MJ/h)	−1527751.5	−108979.5
Flow vapour phase (m³/h)	4659.0	0.0
Flow liquid Phase (m³/h)	0.0	66.4
Flow aqueous phase (m³/h)	0.0	0.0

Composition	% molar	
N_2	0.0069	0.0000
CO_2	0.0086	0.0003
H_2S	0.0000	0.0000
Methane	0.8458	0.0212
Ethane	0.0792	0.0101
Propane	0.0262	0.0079
i-Butane	0.0075	0.0076
n-Butane	0.0089	0.0157
i-Pentane	0.0043	0.0204
n-Pentane	0.0027	0.0206
C_6	0.0021	0.0940
C_7	0.0012	0.1056
C_8	0.0034	0.0816
C_9	0.0000	0.0041
C_{10}	0.0005	0.0930
C_{11}	0.0002	0.0807
C_{12+gaz}	0.0001	0.0533
C_{7-RC}	0.0000	0.0557
C_{8-RC}	0.0002	0.0183
C_{9-RC}	0.0001	0.0185
C_{10-RC}	0.0000	0.0321
C_{11+RC}	0.0000	0.0142
C_{12+oil}	0.0000	0.0122
C_{12+A6}	0.0000	0.1470
C_{12+SE8}	0.0000	0.0674
C_{12+SE9}	0.0000	0.0173
Water	0.0022	0.0013

state and the other in bi-phasic, so the constituents in common are the heavy parts of the vapor, which is the light part of the liquid phase and not condensate.
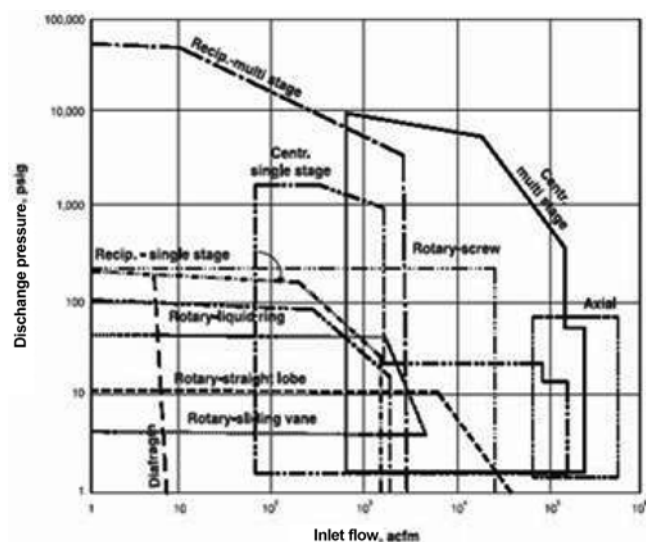
Thus, the liquid part will lose these light elements that will go to the vapor, which lead to an increase in volume of vapor against a

Table 4. Quantities of recoverable gas

			HP separator	MP separator	LP separator	VLP separator	Total
Gas flow-rate (MMscmd)	Phase 1	Summer case	6.809	2.503	0.961	0.314	10.587
		Winter case	6.741	2.441	1.005	0.312	10.499
	Phase 2	Summer case	-	7.065	1.651	0.010	8.725
		Winter case	-	6.981	1.651	0.016	8.648

Table 6. Simulation results without consideration of intermediate supplies

Fluid	Gas product (Kg/h)	Stabilized oil (Kg/h)
Collect HP	241332.0	27600.0
Collect MP	86250.0	10258.0
Collect LP	29323.0	9031.0
Collect VLP	10053.0	8188.0
Total	366958.0	55076.0

Fig. 10. Type of compressor as a function of compressor output pressure and flow of gas to be compressed.

decrease in volume of the liquid phase.

The fact that the liquid phase loses a light element, which will lead to an increase in density, corresponds to a decrease in °API (density is inversely proportional to °API).

CHOICE OF TYPE OF COMPRESSORS

The produced gas in the USC must be compressed before its delivery to the gas treatment unit (GTU) [16].

The delivered pressure from USC must be 89 bars, since the feeding pressure at the GTU is 83 bars, and the pressure loss between USC and GTU is estimated at 6 bars.

The gas HP is sent directly without compression. So, we need three stages of compression: 6 to 20, 20 to 42 and 42 to 89 bars. The choice of type of compressors is made by using Fig. 10 and Table 7, given the type of compressor according to the outlet pres-

Table 7. Gas flow rate for different stages of compression

	Compression from 6 to 20 bars	Compression from 20 to 42 bars	Compression from 42 to 89 bars
Gas flow			
(acmm)	48.174	67.034	156.483
(acfm)	1701.258	2367.298	5526.141

sure and the gas flow rate to be compressed.

COMPRESSION RESULTS

- Compression from 6 to 20 bars can be achieved in a centrifugal compressor, but since the compression ratio is high (3.33), an alternative compressor is more appropriate.

- Compression from 20 to 42 bars is done in a centrifugal compressor.

- The compression from 42 to 89 bars is achieved with a centrifugal compressor.

As the compression ratio from 20 to 42 bars and from 42 to 89 bars is the same (2.1), we can achieve these two compressions using a single two stage compressor.

Hence, the choice of pressure values P2=42 bars and P3=20 bars is made in order to use a single two stage centrifugal compressor instead of two compressors.

The centrifugal compressor is usually driven by a gas turbine. Values of power to be supplied for each compression stage are shown in Table 8.

CONCLUSION

Since the energy of compression is proportional to the flow rate of gas to be compressed, the flow variation curve as a function of pressure should be similar to that of the compression energy variation as a function of pressure. The difference in the shape of curves shown in Fig. 7 and Fig. 8 can be explained by the fact that all feeds HP, MP, LP and VLP have been taken into account for the curve of Fig. 7, whereas only the HP collection has been included for Fig. 8.

This leads us to conclude that intermediate feed stocks in a multi stage separation process affect the shape of the compression energy curve as a function of pressure. The method of minimum compression energy is therefore a graphical method based on the compression energy curve.

The use of intermediate feeds in separation batteries represents a gain in investment (less equipment); however, more gas is produced and the oil quality degraded. An economic study of the scheme should be done to assess its cost effectiveness, in particular when large flows

Table 8. Needed power for each compression

			First stage	Second stage	Third stage	VLP separator
Power (KW)	Phase 1	Summer case	823.964	1822.285	5297.674	7943.923
		Winter case	746.883	1765.804	4597.302	7109.989
	Phase 2	Summer case	25.061	2368.147	12284.657	14677.865
		Winter case	39.450	2182.567	10997.112	13219.129

of oil are to be stabilized and gas produced flared to the atmosphere.

NOMENCLATURE

A   : function of third stage pressure
a, b   : constant Peng Robinson of state equation
acfm   : actual cubic feet per minute
acmm   : actual cubic meter per minute
Gcm   : Giga cubic meter [10^9 m^3]
GOR   : gas oil ratio, scf gas per bbl of oil
HP   : high pressure
MP   : medium pressure
n   : number of stage
K   : constant of phase's equilibrium
LL   : liquid-liquid
LLV   : liquid-liquid-vapour
LP   : low pressure
LV   : liquid-vapour
P   : pressure
P_c : critical pressure
P_1 : pressure of first stage
P_2 : pressure of second stage
P_3 : pressure of third stage
P_4 : pressure of four stage
P_i : pressure of stage (i), initial separator pressure
P_{i-1} : pressure of stage (i−1)
P_n : pressure of stage (n)
R   : gas constant
Scf/bbl : standard cubic feet
T   : temperature
T_c : critical temperature
V   : vbolume
V_c : critical volume
VLP   : very low pressure
Φ : flow
α : Peng Robinson parameter
ω : acentrique factor

Abbreviations

API   : American Petroleum Institute
BICC   : Bureau of Information Climate Change
BWRS   : Benedict-Webb-Rubin-Starling
GTL   : Gas To Liquid
GTU   : Gas Treatment Unit
MEC   : Minimum Energy Compression
NRTL   : Non-Random Two Liquid Equation
PAHs   : Polycyclic Aromatic Hydrocarbons
PFD   : Power Flow Diagram
UNEP   : United Nations for Programme Environment
UNIFAC   : Universal Functional Group Activity Coefficients
UNIQUAC: Universal QUAsi-Chemical
USC   : Unit of Separation and Compression
VOCs   : Volatile Organic Compounds

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