

**An investigation of the optimum separation conditions in the
Degassing stations of one of southern Iraqi oil Field**

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Abstract:

In field separation facilities operation, operators tend to determine the optimum conditions to maximize revenue.

The object of this study is to investigate the present number of separation stages and their optimal conditions for degassing stations of oil Field.

A computer program model was written to predict the optimal conditions for oil field gas-oil separation stations subject to a given crude oil composition , flow rate and feed temperature and pressure using flash calculations with modified Soave-Redlich-Kwong Equation of state.

Nine Bottom-hole well samples of reservoir crude oil was collected and subjected to PVT analysis commonly performed on crude oil, Surface stage separators gas samples at steady condition were taken and analyzed by gas chromatography apparatus to determine the gas composition.

A good agreement was found by comparing theoretical and experimental prediction for gas composition.

The results indicate that the present five separation stages is the optimum number and the values obtained for these optimum stages pressure are (514.7,119.7,42.7,26.7,14.7 psig) for Summer and (464.7,119.7,42.7,26.7,14.7 Psig) for Winter where the current operation pressures are (614.7,119.7,42.7,26.7,14.7 psi) for Summer and Winter. The total liquid yields in stock tank reaches (1.29, 1.31 %) for summer and winter respectively and the stock tank API Gravity reaches (0.41, 0.42) for summer and winter respectively.

Key words: Flash calculation; Multi-stage separation; Optimum separation.

تقييم عدد مراحل العزل الحالية وضغط عزل المرحلة الاولى في محطات عزل الغاز ل احد حقول نفط جنوب العراق

الخلاصة:

في عمليات تشغيل منشآت عزل الغاز الحقلية، يهدف المشغل الى ايجاد ظروف العزل المرهلي المثلى بهدف تعظيم الموارد المالية.

تهدف هذه الدراسة لتقييم عدد مراحل العزل وضغط عزل المرحلة الاولى لمحطة عزل الغاز في احد حقول النفط تحت الدراسة.

تم تهيئة وكتابة برنامج يعمل بالحاسوب لاختيار امثل ظروف العزل لمحطات عزل الغاز عن النفط والتي تخضع لعدة عوامل منها مكونات النفط الخام، معدلات الجريان، ضغط و درجة حرارة النفط الخام المغذي مستخدما حسابات التبخر الوميضي وحل معادلة الحالة لسواف – ريدلج – وونج، تم اخذ عدد 9 نماذج نفط قاع البئر وتحليلها مختبريا، ونماذج غاز من مراحل عزل الغاز عند ظروف العزل المستقر وتحليلها مختبريا لايجاد نسبة مكونات الغاز.

النتائج بينت ان عدد مراحل عزل الغاز الخمسة الحالية (بضمها خزان الجريان) هو العدد الامثل وان امثل ضغوط عزل هي (514.7, 119.7, 42.7, 26.7, 14.7) psig في الصيف و (464.7, 119.7, 42.7, 26.7, 14.7) psig في الشتاء.

وان كمية النفط المنتجة ازدادت بنسبة % (1.29, 1.31) في الصيف والشتاء على التوالي. وان درجة كثافة API ازدادت بنسبة (0.41, 0.42) للصيف والشتاء على التوالي.

Introduction:

The main function of a surface production facility is to separate the well stream into three components, typically called “phases” (oil, gas, and water), and process these phases into some marketable product (s) or dispose of them in an environmentally acceptable manner. In mechanical devices called “separators”, gas is flashed from the liquids and “free water” is separated from the oil. For a given separator, factors that affect separation of liquid and gas phases, including separator operating pressure, temperature, and fluid stream composition. Changes in any of these factors, change the amount of gas and liquid leaving the separator.

In petroleum industry, the optimum pressure is defined as the maximum liquid volume is recovered in the stock tank per volume of reservoir voidage. This pressure corresponding to a maximum in the API gravity, a minimum in total gas oil ratio (GOR),

and minimum oil formation oil factor (B_o) [1]. In general, the optimum separation conditions are fairly complex due to some variables (well stream, flowing temperature and Pressure, liquid content of the residue gas, and composition of the well stream to the separator).

In order to investigate the present stages number and the separation stages pressure for the field are required a knowledge of composition, temperature, pressure and volume relationships of coexisting vapor and liquid phases, phase equilibrium calculations are also needed especially in the vicinity of critical point. These quantitative can be estimated by means of thermodynamic indices, using an equation of state (EOS).

The most widely used equation of state in petroleum industry is the cubic type, such as the Redlich Kwong (RW) equation of state and various modifications [2]. Flash calculations with an equation of state provide the most accurate and reliable method for phase equilibrium production [3]. Modified Soave-Redlich-Kwong (SRK) equation of state has been adopted in this study, and a computer program is developed to perform flash calculations for hydrocarbon and non-hydrocarbon mixture.

Developed a correlation to calculate the optimum second-stage pressure in a three-stage separation system. The inputs required for the calculation are primary pressure; stock-tank pressure; and the mole fractions of methane, ethane, and propane. This correlation does not need flash calculations [4].

Proposed a method to determine the optimum separator pressure for the two-stage separation provided that the stock tank is connected to the atmosphere. In their study, GORs obtained from different separators are plotted against separator pressure, and then the optimum pressure is the pressure that produces minimum GOR [5].

Presented a methodology for optimizing separator pressures in the crude-oil production unit. It can be used to estimate the optimum pressures of separators in different stages of separation. The disadvantage of this method that it requires tremendous numbers of trial-separator pressures and difficult to obtain the exact optimum pressures [6].

Developed a group of correlations for optimum separator pressure for volatile oils using the results of the computer model [7, 8]. These correlations are based on data from

over 6,000 computer model runs with various independent variables. The variables are temperatures of stages, mole fractions of some components of the feed stream, and optimum separator pressures.

Surface Separation:

The crude oil separation is only a part of the entire system. The total system looks very much like that shown in figure (1) which represents a fairly complete processing set up for crude oil and handling gas [9]. The field separation process consists of two or more separators operating in series at lower pressures. Each condition of pressure and temperature at which gas and liquid are separated is called a separation stage. If the pressure of the last separator is greater than atmospheric pressure, the stock tank acts as a stage separation. In each separator, the previous liquid stream is flashed at separator.

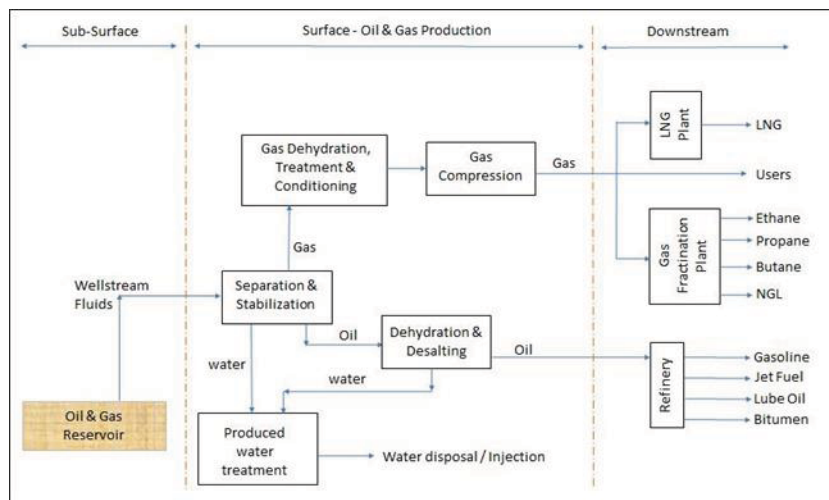


Fig. (1) Typical Oil and Gas Production Schematic

Pressure and Temperature:

The resulting vapor and liquid products are then removed from contact with each other on leaving the separator. Figure (2) shows the five stage separation process system.

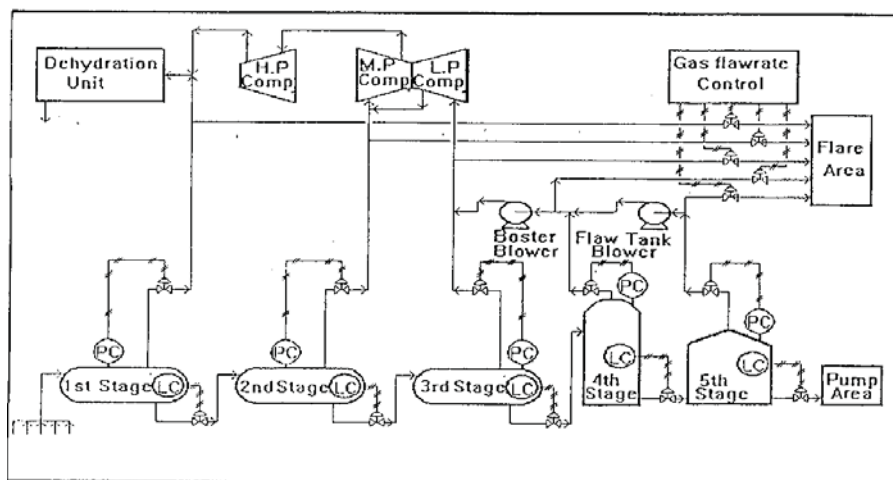


Fig. (2) Gas-Oil Separation Processing Systems.

The feed crude oil is supplied by oil wells gathered at degassing station manifold, and enters the first stage two phase separator in a given composition, pressure, temperature, and flowrate. The pressure and temperature of the entering feed are usually higher than those of the first stage, the pressure drop can cause flash vaporization of some of the gas dissolved in the crude oil.

Gas produced from the first stage, which containing a high percentage of light components such as methane, ethane, propane ... etc., flow to the compression stations. Pressure of the separator is subjected to direct control by means of pressure regulating devices. Crude oil leaving the first stage flow to the second stage which is held at a lower pressure. Again pressure drop causes flash vaporization of the dissolved gas. This process continues up to the final stage which is a stock tank. The stock tank is usually operated at pressure slightly higher than atmospheric.

Under the assumption of equilibrium conditions, and knowing the composition of the fluid stream coming into the separator and the operating pressure and temperature conditions, we could apply our current knowledge of VLE equilibrium (flash calculations) and calculate the vapor and liquid fractions at each stage. However, we are looking at designing and optimizing the separation facility, we would like to know the optimal conditions of pressure and temperature under getting the most economical profit from the operation.

Separator calculations are basically performed to determine [10]:

- Optimum separation conditions: separator pressure and temperature
- Compositions of the separated gas and oil phases
- Oil formation volume factor
- Producing Gas-Oil ratio
- API gravity of the stock tank oil

Constraints:

The objective function for separator pressure optimization can be formulated as follows:

Objective Function = $f(x)$

$$= \begin{cases} \text{Maximize} & \begin{cases} \text{Stock tank oil gravity (API)} \\ \text{Liquid volume in the stock tank } (V_{OST}) \end{cases} \\ \text{Minimize} & \begin{cases} \text{Total Gas Oil Ratio (GOR)} \\ \text{Oil Formation Volume Factor } (B_o) \end{cases} \end{cases}$$

The terms of the stated objective function are equivalent, e.g., if stock tank oil API gravity is maximized, then the oil formation volume factor and gas oil ratio are minimized.

The constraints which are to be considered for the processing system showing in Figure (2) can be expressed as:

1. Stages Number :

$$N \leq N_{Max} \quad \dots 1$$

2. Separator pressure is not exceed the highest allowable pressure specified by the manufacturer:

$$P_i < P_{i \max} \quad \dots 2$$

3. The pressure of a separator couldn't exceed the pressure of the previous separator.

$$P_{i+1} < P_i \quad \dots 3$$

4. The pressure of a separator prior to the stock tank shouldn't be below the stock tank pressure,

$$P_1, P_2, P_3, \dots, P_{N-1} > P_{N-1} \quad \dots 4$$

5. Stock tank pressure is equal to the atmospheric pressure approximately

$$P_N \cong 1 \text{ atm} \quad \dots 5$$

6. Production rate shouldn't exceed a certain value in order to avoid gas slippage and allow good separation:

$$Q < Q_{\max} \quad \dots 6$$

Crude oil specification such as API gravity can be satisfied by blending various cruds having deferent compositions, and called from several oil field, therefore these specifications wouldn't impose as constraints on the gas-oil separation process.

The difficulty for developing a method is to calculate the amount of products for a given set of operating conditions.

In order to solve this problem one has to develop equations for following:

1. Material Balance and Flash vaporization calculations
2. K-values, of various component of crude, as functions of pressure temperature and compositions.

Material Balance and Flash vaporization calculations:

The purpose of the flash calculation on a two-phase system is to establish the amounts of gas and liquid, and the analysis of them. In the usual application, one will need to find the bubble point and dew-point of the system to bracket the condition at two-phase.

The equations needed to make the flash calculation are easily derived by carrying a material balance over the separator. These equations are found in the literature [11] and will not be derived here.

The necessary equations are the following:

$$F \cdot z_i = L \cdot x_i + V \cdot y_i \quad \dots 7$$

$$\sum_i^n x_i = \sum_{i=1}^n \frac{z_i}{1 - \psi(1 - k_i)} \quad \dots 8$$

$$\sum_1^n y_i = \sum_{i=1}^n \frac{z_i \cdot k_i}{1 - \psi(1 - k_i)} \quad \dots 9$$

Where:

$$\psi = \frac{V}{F} \quad \dots 10$$

when each member of equation (8) is summed over all components i and the result so obtained is restated in functional notation, one obtains :

$$f(\psi) = \sum_{i=1}^n \frac{z_i}{1 - \psi(1 - k_i)} - 1 \quad \dots 11$$

Equation (11) can be solved by trial and error, by guessing a value for (ψ) between 0 and 1 until $f(\psi) = 0$.

The most widely employed computer method for solving equation (11) are false position and newton's method [10]. In the latter, a predicted value of ψ root for iteration $k+1$ is computed from the recursive relation:

$$\psi^{k+1} = \psi^k - \frac{f(\psi^k)}{\bar{f}(\psi^k)} \quad \dots 12$$

Where the derivative in equation (11) is

$$\bar{f}(\psi) = \sum_{i=1}^n \frac{z_i(1 - k_i)}{[1 - \psi(1 - k_i)]^2} \quad \dots 13$$

The iteration can be initiated by assuming $\psi^1 = 0.5$.

Sufficient accuracy will be achieved by terminating the iterations when:

$$\frac{\psi^{k+1} - \psi^k}{\psi^{k+1}} \leq 0.0001 \quad \dots 14$$

Equation of State:

Any equation correlating pressure, volume and temperature is called equation of state (EOS). It can be used to calculate gas-liquid equilibrium as an alternative to using K-value correlations. The assumption must be made that the equation of state is predict pressure-volume-temperature relationships for liquids as well as for gases [10].

The K-value needed for flash calculations can be calculated using the modified Soave-Redlich-Kwong equation of state [12] which is recommended for predicting components k-value.

The soave equation is given by equation (15) below:

$$Z^3 - Z^2 + Z(A - B - B^2) - AB = 0 \quad \dots 15$$

The soave equation and its modifications are cubic in compressibility factor, as shown by Edemister [13], equation (15) may yield three real roots but the largest is always taken as vapor compressibility factor and smallest positive root is taken as liquid compressibility factor. It can be solved by iterative method either by using cubic solution procedure or by trial and error technique. The component fugacity in a phase may be calculated once the fugacity coefficient has been evaluated. The relationship between fugacity and fugacity coefficient is given by equation (16) below:

$$\bar{f}_i^N = \phi_i^N x_i P \quad \dots 16$$

In terms of the Soave equation, the fugacity coefficient in the liquid or vapor phase can be calculated from equation (17) once the volume of that phase has been determined.

$$\ln \phi_i^N = \frac{b_i}{b} (z - 1) \ln(Z - B) - \frac{A}{B} \left[\frac{2 \sum x_i \alpha_{ij} a_{ij}}{\alpha a} \right] \ln \left(1 + \frac{B}{Z} \right) \quad \dots 17$$

$$A = \alpha a \frac{PT^2}{R^2} \quad \dots 18$$

$$B = \frac{bPT}{R} \quad \dots 19$$

The equation constant for all pure component are calculated from the critical temperature and pressure and acentric factor. In term of critical constant, a_i and b_i are given by equations (20, 21):

$$a_i = 0.42747 \frac{R^2 T c_i^2}{P c_i} \quad \dots 20$$

$$b_i = 0.08664 \frac{R T c_i}{P c_i} \quad \dots 21$$

For all fluid except Hydrogen, α_i is given by equation (21):

Where

$$\alpha_i = [1 - m_i(1 - \sqrt{Tr_i})]^2 \quad \dots 22$$

Where

$$m_i = 0.48508 + 1.55171\omega_i - 0.15613\omega_i^2 \quad \dots 23$$

The composition averaged parameters αa and b are calculated from equations (24, 25) below:

$$\alpha a = \sum_{j=1}^n \sum_{i=1}^n x_i x_j \alpha_{ij} a_{ij} \quad \dots 24$$

$$b = \sum_{i=1}^n x_i b_i \quad \dots 25$$

The cross mixture parameter $\alpha_{ij} a_{ij}$ is given by equation (26):

$$\alpha_{ij} a_{ij} = (1 - k_{ij}) \sqrt{\alpha_i \alpha_j a_i a_j} \quad \dots 26$$

The binary interaction coefficients k_{ij} are used to improve the vapor-liquid predictions.

A first estimation of un known phase composition must be made using the component k -values calculated by the following empirical equation proposed by Wilson [14].

$$k_i = \frac{1}{Pr_i} \left[\exp \left(5.37(1 + \omega_i) \left(1 - \frac{1}{Tr_i} \right) \right) \right] \quad \dots 26$$

Experimental work:

Bottom hole samples from nine wells were collected and analyzed their composition in a PVT laboratory of Basra oil company. The arithmetic average were taken and assumed to be a representative sample composition of reservoir crudes under study.

Statistical parameter:**Correlation verification and validation:**

Statistical correlation verification and validation are the most important step in the correlation development process. Both quantitative and graphical analyses are used to verify the accuracy of proposed correlations [15].

Graphical error analysis:

Graphical means help in visualizing the accuracy of correlations .

Cross plot Graphic analysis:

In this technique, all the estimates values are plotted against the experimental data, and thus a cross plot is made. A 45 straight line is drawn on the cross plot on which estimated value is equal to experimental value. The plotted data matches the experimental data finally well.

Correlation coefficient:

The correlation coefficient (r) represents the degree of success in reducing the standard deviation by the regression analysis.

It is defined as:

$$r = \sqrt{1 - \left[\frac{\sum_1^n (X_{exp} - X_{est})^2}{\sum_1^n (X_{exp} - \bar{X})^2} \right]} \quad \dots 27$$

Where:

$$\bar{x} = \frac{1}{n} \sum_{i=1}^n (x_{exp})_i \quad \dots 28$$

The correlation coefficient lies between 0 and ± 1 . A value of ± 1 indicates a

perfect correlation, whereas a value of 0 implies no correlation at all among the given independent variables.

Result and discussion:

To determine the optimum stage conditions and the number of stages, a series of flash calculations were performed with various and intermediate stage pressure combinations for four and five separation stages. The results are presented in Table (1), including the optimum number of stages and the stock tank.

API gravity, gas oil ratio and formation volume factor values estimated by the program calculations were plotted as a function of pressure as in Figures (3 A, B, C, D) Figures (4 A, B, C, D), which shows the optimum conditions for summer and winter respectively. It can be seen from Figures (3 D, 4 D) for summer and winter respectively that the stock tank API gravity increase as 4th stage pressure decrease, but the minimum 4th stage pressure couldn't be less than 12 psig because the pressure needed to pushed the oil to the oil tank.

Table (1) Result of four and five separation stages.

| | Stage No. | | | | | API | Total Liquid Yields % |
|---------------|-----------|-------|------|------|------|-------|-----------------------|
| | 1st | 2nd | 3rd | 4th | 5th | | |
| winter | 454.7 | 94.7 | 38.7 | 14.7 | - | 33.97 | 42.272 |
| | 514.7 | 119.7 | 42.7 | 22.7 | 14.7 | 34.24 | 42.751 |
| summer | 464.7 | 94.7 | 34.7 | 14.7 | - | 32.91 | 40.032 |
| | 464.7 | 119.7 | 42.7 | 22.7 | 14.7 | 33.14 | 40.454 |

Field test:

Bank B of 50000 bbl/day design capacity with stock-tank No. 2 of 70000 bbl/day flowing capacity in a degassing station of the field under study was chosen for carrying out the field tests. The tests were conducted under the same existing and predicting optimum pressure sets. During the test, stock tank API gravity, separators temperature and pressure, Samples of gas outlet of each separation stage (including flow tank) at steady separation conditions were evaluated.

To avoid turbulence, agitation and undesirable carryover of oil particles with gas, the liquid level must be kept at minimum set. Table (2) shows that the oil level of stage separation versus stage number.

Table (2) Stage separation oil level %

| Stage No. | Stage separation oil level % | |
|-----------------|------------------------------|--------|
| | Summer | Winter |
| 1 st | 35 | 40 |
| 2 nd | 35 | 35 |
| 3 rd | 35 | 35 |
| 4 th | 35 | 35 |

The results of carry over tests using spot test on a white paper were carried out during the field test. It was found that no carry over was taking place at first and other stages due to reduction of the first stage pressure.

The use of SRK equation of state for determination of theoretical oil and gas of individual compositions at stages separation conditions have shown that the theoretical calculated values fit well within acceptable limits with field experimental test results determined from actual measurement using Gas Chromatography tests. This is shown exclusively on Figure (5 A, B) for summer and winter respectively . The plotted points of this study's correlation fall very close to the perfect correlation of the 45° line.

Figure (6) summarizes the correlation coefficient of all separation stages for summer and winter respectively, which indicate the model is sufficient to describe the data.

It was found from Table (3) that the optimal number of stages was five plus the stock tank including the five stage and actual and recommended optimal operation pressure for each separating stage.

Table (3) Operating and recommended optimum operation pressure for each separation stage.

| Stage No. | Present Operating Stage Pressure (Psig) | | Recommended Optimum Stage Pressure (Psig) | |
|---------------------------------------|---|--------|---|--------|
| | Summer | Winter | Summer | Winter |
| 1st | 614.7 | 614.7 | 514.7 | 464.7 |
| 2nd | 119.7 | 119.7 | 119.7 | 119.7 |
| 3rd | 42.7 | 42.7 | 42.7 | 42.7 |
| 4th | 26.7 | 26.7 | 26.7 | 26.7 |
| 5 th | 14.7 | 14.7 | 14.7 | 14.7 |
| Stock tank API Gravity | 32.73 | 33.82 | 33.14 | 34.24 |
| Total liquid yield in Stock tank % | 40.16 | 42.44 | 41.45 | 43.75 |

Conclusions:

- Calculated flash values following the SRK equation based on a composition mixture have shown a good agreement with measured values.
- Optimization of separation pressure for the prevailing oil specifications shown that in order to achieve efficient gas separation or to produce a high quality oil at the stock tank the first stage pressure had to be reduced.
- At the maximum rate of 70000 bbl./day , the banks have to be operated with oil levels cut down to nearly third in the first separators stage in order to bring gas velocity down to acceptable limits and prevent excessive oil carry over with the gas.

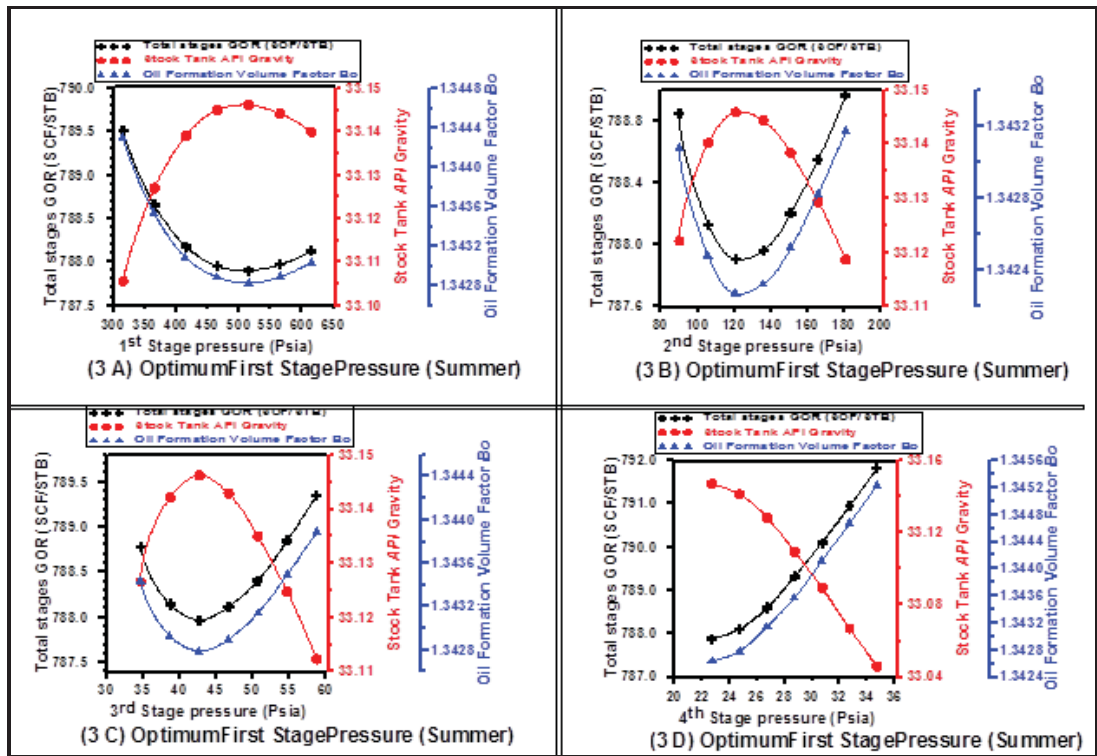


Fig. (3) Optimum first stage pressure (Summer)

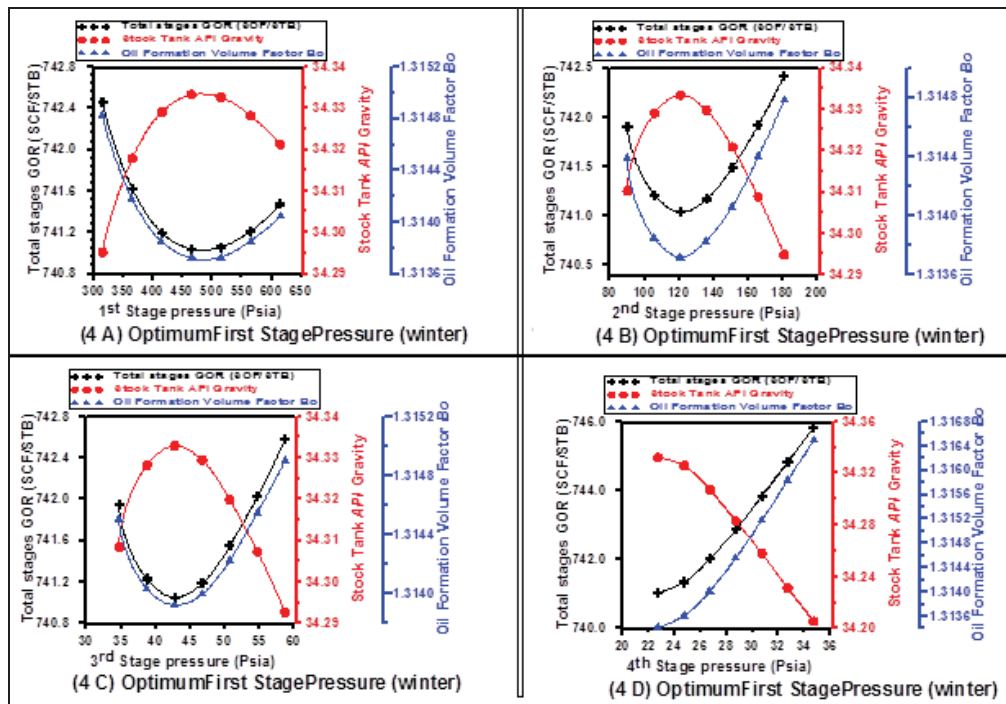


Fig. (4) Optimum first stage pressure (Winter)

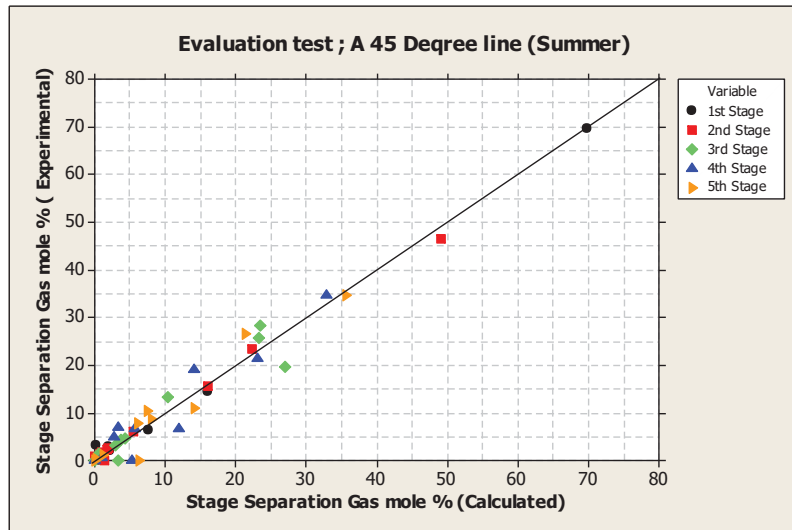


Fig. (5A) Comparison between Experimental and Estimated gas mole fraction of individual component at all five stages (Summer).

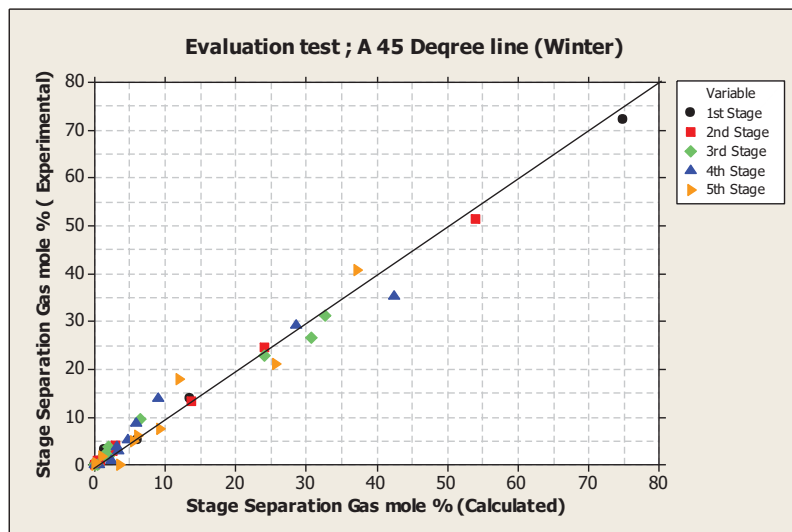


Fig. (5B) Comparison between Experimental and Estimated gas mole fraction of individual component at all five stages (Winter).

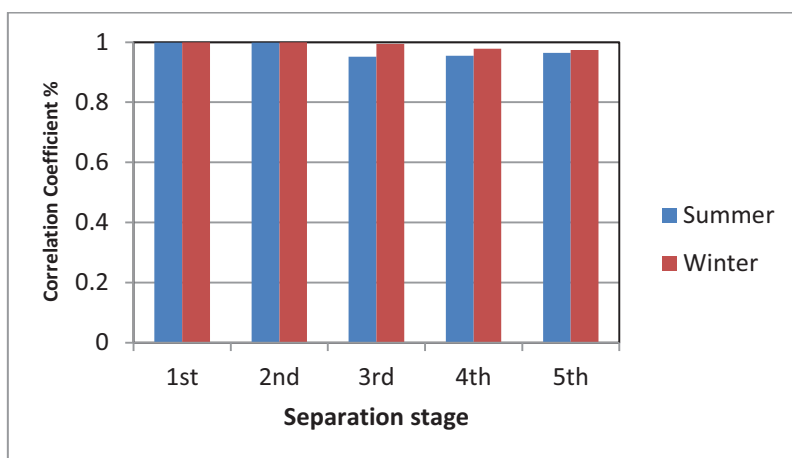


Fig. (6) Evaluation test: Experimental vs. Estimated

Nomenclature:

| Symbols | |
|-----------------------------|---|
| A_{ij} | Constant in Soave-Redlich-Kowing |
| $a, (a_1, a_2, \dots, a_n)$ | |
| B, b, b_i | |
| F | Total moles of feed entering separation process |
| f | Fugacity |
| GOR | Gas Oil Ratio |
| k_i | Vapor liquid equilibrium ratio (K-value) |
| K_{ij} | Binary interaction coefficient |
| L | Total moles of liquid phase leaving separation process |
| M_i | Constant in Soave-Redlich-Kwong |
| M_w | Molecular weight |
| n | Number of component |
| P | Pressure |
| P_{c_i} | Critical pressure of a component |
| P_{r_i} | Reduced pressure |
| R | Gas constant |
| Sp. G | Specific gravity |
| T | Temperature |
| T_{c_i} | Critical temperature of a component |
| T_{r_i} | Reduced Temperature |
| T_o | Reference temperature |
| V | Total moles of vapor phase leaving separation process |
| X_i | Mole fraction in liquid phase |
| Y_i | Mole fraction in vapor phase |
| Z | Compressibility factor |
| Z_i | Mole fraction in mixture phase |
| Greek Letter | |
| β_i | Equation of state parameter |
| α_i | Equation of state parameter |
| ρ | Density |
| ϕ_i^V | Fugacity coefficient of component in a mixture vapor phase |
| ϕ_i^L | Fugacity coefficient of component in a mixture liquid phase |
| w_i | Acentric factor |
| ψ | Mole ratio of vapor to feed |

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