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Economic Evaluation of an Associated Gas Production Facility: A Case Study of X Field in Southern Nigeria

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Abstract— *A surface production facility is an integral part of a gas oil separation process (GOSP) plant. In this study, an economic evaluation of two field handling alternatives for associated gas recovery in an integrated surface production facility located in X Field in Southern Nigeria is performed. The process alternatives evaluated were a four-stage separation facility with four compressors and a four-stage separation facility with three compressors, respectively. The former is the baseline facility at the X Field while the latter is an optimized version of the baseline facility. The economic assessments of the two alternatives were compared. Depending on the alternatives, the crude oil and associated gas facility could produce up to 48,470 to 50,090 bbl/day of crude oil and 50.27 to 50.48 MMSCFD of associated gas. The capital costs (CAPEX), operating costs (OPEX), revenue (REV), and energy demand (ED) for the baseline scenario and its optimized alternative were calculated. The CAPEX for the baseline and optimized scenarios are approximately 24,701,200 USD and 17,491,800 USD, respectively, while the OPEX are approximately 3,859,240 USD and 3,857,900 USD, respectively. Judging by the results,, the optimized alternative is economically more feasible since its evaluation in terms of net profit over a twenty-year period, shows an increase from 21,495.02 MMUSD to 22,141.44 MMUSD, a net profit of 646.42 MMUSD. Further analysis indicates that the payout time shows a differing trend from baseline values, and also gives different outcomes when considering total annualized cost (TAC). Overall, the results show that the optimized process scheme provides better economic performance relative to the baseline scheme under similar feed compositions.*

Keywords— *Crude oil recovery, associated gas recovery, economic analysis, surface production optimization.*

I. INTRODUCTION

At the surface production facility located at the X Field in Southern Nigeria, reservoir fluids are separated as they pass through staged gas-liquid separators operated at consecutively lower pressures. In order to maximize fluid recoveries and minimize operating costs at the X-Field plant, a plausible action would be to determine means to optimize the surface production facility while simultaneously maintaining an optimum separator operating conditions [1, 3, 4, 5, 6, 12]. One optimization strategy is a reduction in the number of process equipment in relevant units of the plant [3]. This X-Field, which is operated in a swamp, consists of flowing wells (and some shut-in wells) at either high pressure, medium pressure or low pressure, depending on their Shut-in Tubing Head Pressure (STHP) [3], The objective of this study is to economically evaluate the current plant (baseline) standards relative to the economic performance of its optimized version, using indicators like capital and operating cost, gross and net profit, net present value, total annualized cost (TAC), utility cost, and compressor power demand.

II. PROCESS DESCRIPTION

Two types of crude oil and associated gas recovery schemes were considered in this study. The first is the Base Case Process (BCP), and the second is the Optimized Case Process (OCP). The BCP is the baseline process of associated gas recovery, from which the OCP scheme emerged. Both selected process models were simulated using a commercial software [3, 14].

A. Base Case Process (BCS)

The schematic diagram of BCS scheme is depicted in Fig. 1. Downstream of a Christmas tree, the produced associated gas, at a relatively high pressure, enters the first high pressure two-phase separator, where it is separated into liquid and gas phases. The gas phase which is within the gravity settling section of the separator flows through the mist extractor, at the top, to the high pressure compressor (HPC). The crude oil collected at the lower section of the separator flows through the bottom exit to the second medium-pressure separator, where both liquid and gas are further separated (or recovered). The third and final low pressure separator receives the liquid bottoms from the second separator for further recovery before reaching the stock tank. For this baseline process, four compressors, including high-pressure compressor (HPC), medium-pressure compressor (MPC), first low-pressure compressor (LPC1), and second low-pressure compressor (LPC2), which are obviously of varying operating pressures are in use with attendant cost implications.

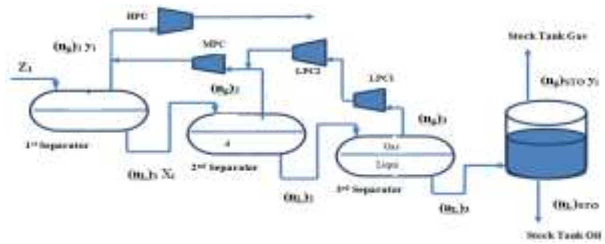

Figure 1. Simplified schematic of the modeled X Field's surface production facility [3].

B. Optimized Case Process (OCP)

OCP is the optimized case process scheme that uses just three compressors for same BCS operation with the objective

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to improve economic performance of the surface production facility by way of recovering more of the different fluids with less compression work. Figure 2 shows the schematic of the OCP scheme. Here, the recovered gas from the second separator is no longer connected to the two compressors. This time, stream 1 from second separator is first sent to the stream entering the first high pressure compressor, which is quite different from what happens in the baseline case (BCS), where the exit gas from the third separator connects with the stream coming from the second separator. Exit stream from third separator then moves to the stock tank and is thereafter transported out of the surface production facility.

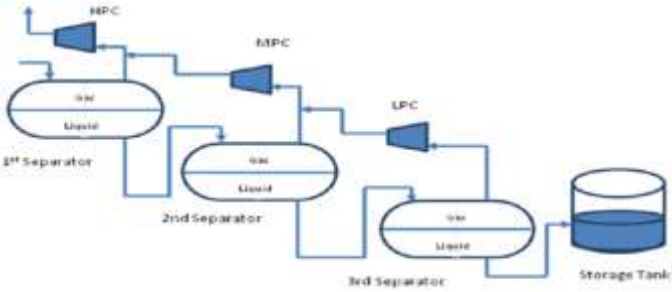

Figure 2. Simplified schematic of the optimized X-Field's surface production facility [3].

C. *Fluid Composition and Specifications*

The data used for this study was obtained from the X Field facility, in Southern Nigeria. Some of the data include: detailed process flow diagram of the surface production facility, inlet feed operating parameters, comprehensive crude oil compositions, equipment summary, and other relevant unit operating conditions and utilities. Table 1 lists the composition of the inlet feed stream to the surface production facility.

TABLE 1. Composition of the Inlet feed stream to surface production facility (SPF) [3].

Composition	Mol. Fraction (%)
N2	0.91
CO2	0.16
C1	34.47
C2	9.67
C3	6.95
i-C4	1.44
n-C4	3.93
i-C5	1.44
n-C5	1.41
n-C6	4.33
C7+	35.29

D. *Methods*

A commercial software [14] is used for simulation work. For the simulation, natural gas components were characterized as either pure or pseudo-components. The pseudo-components were created and added prior to simulation. We also considered the use of Peng-Robinson fluid package, particularly because of the presence of polar and non-polar hydrocarbons, as well as a three-phase system, and it was possible to estimate the binary interaction and activity of the components in both the liquid and vapour phases. The general form of the equations based on the Peng-Robinson computation, including its coefficients are expressed as follows [7]:

$$P = \frac{RT}{v-b} - \frac{a(T)}{v(v+b)+b(v-b)} \tag{1}$$

$$R = 8.314472\,\frac{J}{mole.K} \tag{2}$$

$$b_i = 0.07779608\,\frac{RT_{cr,i}}{P_{cr,i}} \tag{3}$$

$$a_i = 0.45723552\,\frac{R^2\,T_{cr,i}^2}{P_{cr,i}} * \alpha_i(T) \tag{4}$$

where $T_{cr,I}$ and $P_{cr,i}$ are the critical temperature and pressure of each component, respectively, $\alpha(T)$ is the alpha function expressing the temperature dependency of a_i. The alpha function is expressed as follows:

$$\alpha_i(T) = \left[1 + k\left(1 - T_{r,i}^{1/2}\right)\right]^2 \tag{5}$$

where k is,

$$k = 0.37464 + 1.54226w_i - 0.26992w_i^2 \tag{6}$$

$T_{r,j}$ and w_i are the reduced temperature and acentric factor of component i, respectively. This unique alpha fuction was first introduced by Redlich and Kwong [8] and Georgio Soave upgraded the Redlich and Kwong's function as indicated by equation (3). The coefficients a and b in Equations (3) and (4) are determined following mixing rule as:

$$a = \sum_{m=1}^{c}\sum_{n=1}^{c} y_n y_m \sqrt{\alpha_n \alpha_m}\left(1 - \mu_{m.n}(T)\right) \tag{7}$$

$$b = \sum_{n=1}^{c} y_n b_n \tag{8}$$

In this work, the K-values are calculated based on the following:

$$K_{ji} = \frac{\hat{\varphi}_{ji}^{L}}{\hat{\varphi}_{ji}^{V}} \tag{9}$$

where $\hat{\varphi}_{ji}^{L}$ and $\hat{\varphi}_{ji}^{V}$ (expressed in equations (10) and (11)) are the fugacity coefficient of liquid and vapor phases for component "i" in the mixture, at the temperature and pressure of tray "j", respectively:

$$\hat{\varphi}_{ji}^{L} = \frac{\hat{f}_{ji}^{L}}{Px_{ji}} \tag{10}$$

$$\hat{\varphi}_{ji}^{V} = \frac{\hat{f}_{ji}^{V}}{Py_{ji}} \tag{11}$$

where f_i^L and f_i^V are the fugacity of component "i" in liquid and vapor phase, respectively.

E. *Process Optimization*

Our approach to optimization is to consider the total compression unit and its power consumption. We reduced the number of compressors by one, with the objective to reduce the total compressor horsepower. The compressor power demand is estimated using equation 12 [15],

$$H_{pg} = \frac{q_1 p_1}{229E_p}\left(\frac{Z_1+Z_2}{2Z_1}\right)\left(\frac{r^{R_p}-1}{R_p}\right) \tag{12}$$

where H_{pg} is compressor power demand (hp), $P1$, $P2$ and $T1$ are the compressor suction pressure (psia), compressor discharge pressure (psia), and compressor inlet temperature (Fahrenheit), respectively, $Z1$ and $Z2$ are compressibility factor at inlet and outlet, respectively.

The annual energy cost is expressed as [15, 16],

$$\text{Annual Energy Cost } (\$) = H_{pg} \times EE_{cu} \times t \quad (13)$$

where; EE_{cu} is the electrical energy cost per unit ($/kWh) and t is the compressor operating time (*hours*). From equation 13, cost savings through the minimization of compressor power demand can be calculated. Computation of compressor power consumption can be done using the equation 14 [15, 16],

$$\frac{power}{stage} = \frac{m(h_2 - h_1)}{Efficiency} \quad (14)$$

where m is the mass flow rate, h_2 and h_1 are the specific enthalpies at suction and discharge of the compressor, respectively. For an isentropic (reversible and adiabatic) process, the power requirement is given as [15, 16],

$$\frac{Power}{Stage} = \frac{k}{k-1}\frac{T_1 Z_a}{\eta} q \frac{P_s}{T_s}\left[\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}}\right] \quad (15)$$

The following power equation [17] can also be used for an isentropic process,

$$Power = \frac{k}{k-1} T_1 Z_a m R \left[\left(\frac{P_2}{P_1}\right)^{\frac{n-1}{n}} - 1\right] \quad (16)$$

while for a polytropic process, the power requirement is given as [17],

$$\frac{Power}{Stage} = \frac{k}{k-1}\frac{T_1 Z_a}{\eta} q \frac{P_s}{T_s}\left[\left(\frac{P_2}{P_1}\right)^{\frac{n-1}{n}} - 1\right] \quad (17)$$

Using the above equations, the power requirement for the centrifugal compressors in the X Field plant can be estimated if the inlet and outlet conditions of temperature and pressure as well as the composition of the fluid being compressed are known.

III. Results from a Comparison of the BCP And OCP Processes

The results of economic evaluation for the two schemes is shown in this section for good comparison. The feed stream was defined, with flow rate, temperature and pressure conditions given as 104 MMScfD, 41.2 ^oC, and 52 bar, respectively. Economic evaluation is performed using theoretical analysis and a commercial simulator [14]. The economic evaluation entails sizing and mapping both BCP and OCP alternative processes to ensure consistency in its results. The discounted cash flow (DCF) analysis - a wide economic assessment method, is used in estimating the profitability of each plant. Total profit, future value (FV), net present value (NPV), and profitability index (P.I) are estimated as follows [10, 11, 13],

$$Total\ Profit = Total\ Revenue - Total\ Cost \quad (18)$$

$$FV = PV \times FVIF = PV \times (1+r)^t \quad (19)$$

$$NPV = \sum_{t=1}^{k} \frac{NCF}{(1+r)^t} \quad (20)$$

$$P.I = 1 + PVR = 1 + \frac{NPV}{PV_{of\ CAPEX}} \quad (21)$$

where *FVIF* is the future value interest factor, *r* is the nominal interest rate, *t* is the time period, and *NCF* is the nominal cash flow.

A. *Capital and Operating Cost*

The total capital cost result shows that BCS scheme has the higher value, whereas the OCP scheme has the lower value due to its simpler configuration and fewer number of compressors. The total operating cost is mainly composed of the utility cost and raw material cost. The OCP process scheme has the lower raw material cost compared to the baseline scheme, in addition, it has the lower capital cost. The raw material saving for the scheme OCP is better.

Because this study focused on same feed conditions, the higher efficiency of the OCP scheme is expected. In addition, utility costs like electricity are the main utilities and directly influenced by the cost of fuel. The cost of utilities, such as electricity and refrigeration, and the feed raw material costs are the main operating costs considered in this work. The raw material cost here represents the cost of feed natural gas. Table 2 compares the utility costs of the BCP and OCP schemes.

According to the economic analysis shown in Table 2, the OCP scheme has the most favorable utility cost. However, the BCP scheme gives better results for cooling water cost, while the OCP scheme compression has the lower total capital cost due to its lesser number of equipment. The results indicates that that using the OCP scheme would save utility costs of approximately 10.3 United States Dollars per hour.

TABLE 2. Comparison of some utility cost for the BCP scheme (Existing) and OCP scheme (optimized).

Utilities	BCP		OCP	
	Rate	Cost per hour (USD/hr)	Rate	Cost per hour (USD/hr)
Electricity	1938.76 KW	150.254	1820.67 KW	141.102
Cooling Water	0.02273MMGal	2.72748	0.03747MMGal	4.49616
Refrigerant (Feron12)	59.2149 KLB	5.03327	24.5504 KLB	2.08679
Total cost (USD/Hr)		158.015		147.685
Amount saved due to Optimization, USD/hr			10.33	

Table 3 is an investment summary of the BCP and OCP plants, indicating that a total operating cost of 601,340 USD can be saved if the OCP scheme is used.

TABLE 3. Investment summary of the BCP scheme (Existing) and OCP scheme (optimized).

	BCP Plant	OCP Plant
Total capital cost (USD)	24,701,200	17,491,800
Total operating cost (USD)	3,859,240	3,257900

A profitability analysis is conducted for each scheme in order to compare the net profit. The net profit is obtained by gross profit minus an assumed tax rate. The gross profit is calculated by products revenue minus operating cost [10]. The gross profit of the BCP and OCP schemes are different with the OCP schme relatively higher than the BCP scheme (see Table 4), such that increase in net profit over a 20 year period is about 646.42 MMUSD.

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TABLE 4. Total Gross profit and total net profit from liquid production from the BCP (existing) and OCP (optimized) plants over a 20-year period.

	BCP Plant	OCP Plant
Total Gross Profit (MMUSD)	30,707.16	31,630.63
Total Net Profit (MMUSD)	21,495.02	22,141.44

The total annualized cost is the annualized value of the total net present cost. The total annualized cost calculations can be done using equation (22). Here, the TAC is defined as the sum of annualized TCC and, total operating cost minus byproduct credits [10]. Thus, we defined the total annualized cost (TAC) as,

$$TAC=ATCC+TOC-BPC \qquad (22)$$

where ATCC is annualized total capital cost, TOC is total operating cost, and BPC is the byproducts credits. The TAC value for each feed is shown in Fig. 3.

For same feed, the OCP (optimized) scheme had the lower TAC value, when compared to the BCP (existing) scheme.

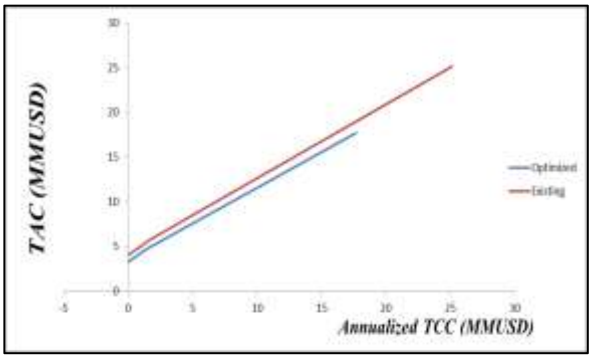

Figure 3. TAC analysis.

B. Crude Oil and Associated Gas Production

Table 5 shows the behavior of the two schemes in terms of production of crude oil and associated gas. We observed a crude oil and natural gas production increase of 1620 bbl and 210 MSCF, respectively.

TABLE 5. Gross liquid and gas production from the existing (BCP) and optimized (OCP) plant

	BCP Plant	OCP Plant
Gross liquid production (bbl./day)	48,470	50,090
Gas production (MMSCF/D)	50.27	50.48

C. Net Present Value

Generally, the present value of a project is simply the sum of the present values of all individual annual net cash flows over the life time of a project [10]. The Net Present Value (NPV) recognizes the net present value of money and applies equal weight to all future incomes. Table 6 indicates an increase by 122.08 MMUSD in the sum of the present value of individual annual cash flows of the optimized plant (OCP scheme) over the existing plant (BCP scheme) during the duration of the project which is a 20-year period.

TABLE 6. Net Present Value from liquid production from the existing and optimized plant over a period of 20yrs

	BCP Plant	OCP Plant
Net Present Value (MMUSD)	3,872.82	3,994.90

D. Compressor Power Demand

Having performed an economic evaluation of the X-Field plant, we can now analyze the performance of the compression system using both correlations and a commercial simulator. The performance of the plants is summarized in Table 7.

The total compressor power demand based on the use theoretical correlations for the BCP and OCP schemes are 5416 and 5143 hp, respectively. The corresponding values from using the commercial simulator are 4167.6 hp (BCP scheme) and 3941.3 hp (OCP scheme), respectively. The use of either computation method will result to a saving of compressor horse power, as demonstrated by the 273 hp and 226.3 hp values. These results clearly suggests that the efficiency of the compression system is relatively higher with the OCP scheme, regardless of the method used in computing the compressor horsepower.

TABLE 7. Compressor power demand for existing (BCP scheme) and optimized (OCP scheme) Plants.

	Compressor power demand for existing (BCP) plant (hp)		Compressor power demand for optimized (OCP) plant (hp)	
	Theoretical Correlation	**Software Result**	**Theoretical Correlation**	**Software Result**
LPC 1	281.0	241.9	580	488.3
LPC 2	218.0	175.7	0.0	0.0
MPC	1444.0	1122.0	1704.0	1346.0
HPC	3473.0	2628.0	2859.0	2097.0
Total power consumption	5416.0	4167.6	5143.0	3941.3
Computed/saved Power (correlations), Hp			273.0	
Computed/saved Power (software), Hp			226.3	

IV. CONCLUSION

In this work, key economic indicators were used to compare the overall economic performance of the crude oil and associated gas recovery schemes (the BCP and OCP schemes, respectively). The economic evaluation included economic indicators such as the annualized total capital cost, operating cost, net profit, net present value, and compressor power demand. The results show that the OCP scheme has the better performance with same feed conditions. Further, the OCP scheme has the lower capital cost due to relatively less configurational complexity, and also gives the better performance in terms of minimum raw material cost with respect to its higher efficiency, and lower compressor energy demand.

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