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Development of a Predictive Model for Estimating Glycol Loss by Vaporization in TEG Dehydration of Natural Gas

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Abstract
Control of glycol losses has remained a major challenge in Triethylene Glycol (TEG) dehydration of natural gas. Fugitive and carryover losses are largely minimized by applying suitable engineering coupled with meticulous unit operations. But vaporization losses are unpreventable as a result of the vapour pressure of TEG with respect to the dehydration process conditions. This work presents a model for estimating glycol losses by vaporization. The TEG dehydration process was simulated for glycol losses using HYSYS software with its Peng-Robinson Equation of State (PR-EOS). The key performance indicators used in the simulation include: gas flow-rate, gas temperature, contactor pressure, lean glycol concentration, lean glycol circulation rate, lean glycol temperature, reboiler temperature, reboiler pressure, stripping gas rate, regenerator reflux flow rate and the number of trays. The gas composition is as is common in Nigerian Niger Delta region. The result of the simulation provided the input data for a 12-Run Plackett-Burman Design of Experiments and subsequent multiple linear regression. The outcome was a linear model with coefficient of determination (squared correlation coefficient), $R^2 = 1$.

Keywords: *TEG, Dehydration, Vaporization, Simulation, Design of Experiment, Modeling*

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I. INTRODUCTION

In natural gas conditioning, TEG dehydrators account for over ninety percent of the natural gas dehydration facilities in operation. Moreover, about sixty percent of the operating costs of these glycol units is attributable to glycol replenishment and replacement [6]. This has imposed a demand for improvements in efficiency and reductions in operating expenses of glycol units as the most imperative requirement for the economic operation of a glycol dehydration unit.

There are several factors which cause glycol losses in process, including carryover and the excessive foaming in the absorber, glycol degradation due to ageing, as well as due to technical problems associated with equipment corrosion and inappropriate absorber temperature and pressure. These losses are categorized as either mechanical (as a result of leakages and spillages), carryover from the contactor (from design and /or operational faults) or as vaporization from the contactor and regenerator (owing to the physical properties of glycol and from operational errors). However, with the exception of evaporation losses, considerably effective measures can be put in place to mitigate losses originating from mechanical and fugitive sources.

Evaporation losses are unavoidable as a result of the combined effects of the dehydration process conditions and the physical properties of glycols, particularly the vapour pressure. The glycol contactor (high pressure, low temperature), and regenerator (high temperature, low pressure) are the most significant sections in the dehydration system where most of the glycol losses by vaporization occur. Though glycol vapor pressures are very low, some very little concentration of glycols may always end up in the treated gas phase. Losses by entrainment and foaming will further increase the content. They are therefore the key areas of attention when considering the phenomenon of glycol consumption by way of evaporation.

The common assumption is that vaporization losses are insignificant [8], and this could probably be the reason for the paucity of research efforts to quantify it as is the case with fugitive and carry-over losses. This work focuses on developing a predictive model for estimating glycol losses by vaporization as a step towards proper identification and management of this category of losses.

1.1 TEG Dehydration Process Description

The TEG dehydration process is dividable into two main parts: gas dehydration and solvent regeneration. In the dehydration stage, water is removed from the gas; while in the regeneration stage, water is removed from the solvent. After this process, the fresh glycol solution is ready again to be used in the contactor tower. Fig.1 is a typical process flow and equipment schematic diagram of natural gas dehydration using TEG [3]. While the overall process equipment is similar for all glycol dehydration units, there can be considerable variations among installations.

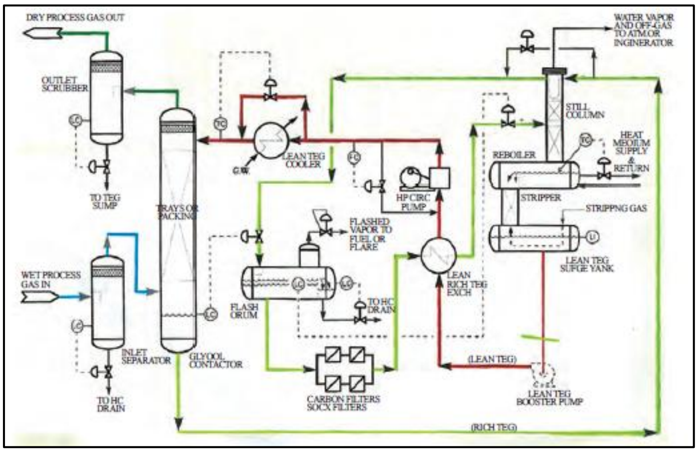


Figure 1: Typical PFD for TEG Dehydration [3]

The wet gas flows through a separator to get rid of condensed liquids or any solids that might be in the gas [2].

Some absorbers incorporate the separator in a bottom section of the vessel, in which case the gas then flows upward through a chimney tray into the glycol absorber portion of the vessel. The glycol contactor or absorber can contain trays, random packing, or structured packing. A trayed contactor is usually fitted with several bubble-cap trays. Lean glycol is pumped into the upper portion of the contactor, above the top tray but below a mist eliminator. The trays are flooded with glycol flowing down from tray to tray through the downcomer sections. The wet gas rises in counter-current flow through the bubble caps and is dispersed as bubbles through the glycol on the trays. This provides an intimate contact between the gas and the glycol. The glycol being naturally highly hygroscopic, most of the water vapor in the gas is absorbed by the glycol. The rich glycol, containing the absorbed water, is withdrawn from the contactor near the bottom of the column above the chimney tray through a liquid level control valve to the regeneration section. The treated gas passes through a mist eliminator located at the overhead section of the column and out usually meeting the specified water content.

The rich glycol is passed through a heat exchange coil in the top of the reboiler column called the still. The heat exchange generates some reflux for the separation of the water from the glycol in the top of the still and also heats the rich glycol some degrees. In some installations, the rich solution passes to a flash tank operating at about 15 to 50 psig, which allows absorbed hydrocarbon gases to separate from the glycol. The glycol then flows into the still through some sock and/or carbon filter beds to a heat exchanger, exchanging heat with the regenerated glycol. It gravitates through a packed section in the still into the glycol reboiler section, where it is heated to the control temperature for regeneration at close to atmospheric pressure. At this high temperature, the water evaporates and leaves through the top of the still. The regenerated glycol flows into the surge tank, from where it goes through the lean/rich heat exchanger to the glycol booster pumps. The pump boosts the flow of the lean glycol to the contactor pressure. Prior to entering the contactor, it exchanges heat with the dry gas leaving the contactor or some other heat exchange medium in order to maintain a temperature difference between the lean glycol over the dry gas of $5^{O}C$ to $10^{O}C$ [9]. High lean TEG temperature increases the moisture content of the outlet gas as it reduces capacity of TEG absorb moisture. On the contrary, lower temperatures increase glycol loss as a result of foaming while boosting the absorption of BTEX hydrocarbons which constitute environmental hazard if discharged to the atmosphere [10].

II. DATA ACQUISITION

Process plant data acquired from the Daily Report for Gas Dehydration, the Flash Daily Report in X-Gas Plant (located in Nigeria's Niger Delta) and from extant literature (Arnold and Stewart,1989; Campbell, 2002; Dow Chemical Company, 2010; GPSA Data Book, 1998; Ikoku, 1992; Kidnay and Parrish, 2006; Dow Chemical Company, 2003; Maddox et al, 1992) were used. The independent variables from the Ob-Ob Plant Daily Reports include the following: gas flow-rate, gas temperature, contactor pressure, lean glycol concentration, lean glycol circulation rate, lean glycol temperature and the process gas composition which is as is common in Nigerian Niger Delta (Table 1). From literature, the following were included: reboiler temperature, reboiler pressure, glycol circulation ratio, stripping gas rate, and regenerator reflux flow rate (Arnold and Stewart, 1989). The minimum and maximum values have also been fixed for the application of the Placket- Burman Design of Experiment. These are presented in Table 1.1.

2.1 Key Performance Indicators

The key independent process variables which affect glycol vaporization losses are presented in table 2. Their average maximum and minimum values are also indicated. The equilibrium at the top of the absorber depends on the glycol circulation rate and the number of trays/stages of packing. The reboiler temperature in the regenerator and the amount of stripping gas used (if it is used) determine the equilibrium water content, because they limit the purity of the lean glycol to the absorber. The operating pressure of the regenerator affects the lean glycol purity as well. The temperature of the inlet gas stream controls the total amount of water to be removed; lower temperatures mean that less water is absorbed by the glycol. Also the lean glycol temperature at the top of the absorber affects the water partial pressure at the top equilibrium stage, which means that high glycol temperatures lead to large water content in the outlet gas. The temperature difference between the rich glycol and the dry gas has consequences for glycol losses.

Table 1: Process Gas Composition of X-Gas Plant

	Inlet Gas, %Mol	**Outlet Gas, %Mol**
Methane C_1	81.46	81.46
Ethane C_2	8.85	8.86
Propane C_3	4.52	4.53
Iso-Butane iC_4	0.85	0.85
Butane C_4	0.94	0.94
Pentane C_5+	0.88	0.88
CO_2	2.21	
N_2	0.12	
Water H_2O	Saturated 96.14 lbs/MMSCF	≤ 4.7 lbs/MMSCF

Table 2: Key Performance Indicators that impact Vaporization Losses

S/N	**PROCESS VARIABLE**		**UNITS**	**MIN**	**MAX**
1	Gas Flow-rate	Q_{Gas}	Std m^3/hr	200	400
2	Gas Temperature	T_{Gas}	^OC	25	40
3	Contactor Pressure	P_{Contac}	Bars	34	83
4	Lean Glycol Concentration	cTEG	%	95	99,7
5	Lean Glycol Circulation Rate	Q_{TEG}	m^3/hr	6	25
6	Lean Glycol Temperature	T_{TEG}	^OC	10	60
7	Reboiler Temperature	T_R	^OC	182	204
8	Reboiler Pressure	P_R	Bars	0.1	0.6
9	Number of Trays	N	-	6	12
10	Stripping Gas Rate	Q_{strip}	Std m^3/m^3TEG	0	90
11	Condenser Temperature	$T_{cond.}$	^OC	0.2	1.0

III. HYSYS SIMULATION

Hysys software with its Peng-Robinson Equation of State was used to simulate the natural gas dehydration process using Triethylene glycol (TEG) to determine the glycol loss. The flow diagram is shown in Figure 3.5. The water-saturated gas with a water content of 96.14 lb/MMSCF enters the bottom of the contactor column at 40°C (100°F) and 8300 kPa (1204 psia) at a rate of 200 std m^3/hr (3365 kgmole/hr). The contactor column has six theoretical trays. The lean TEG solution with a concentration of 99.7 and temperature of 10^0C enters at the top of the contactor column and flows down in the column. An overview of the contactor as well as the contactor worksheet is shown in Figure 2 and Table 3 respectively.

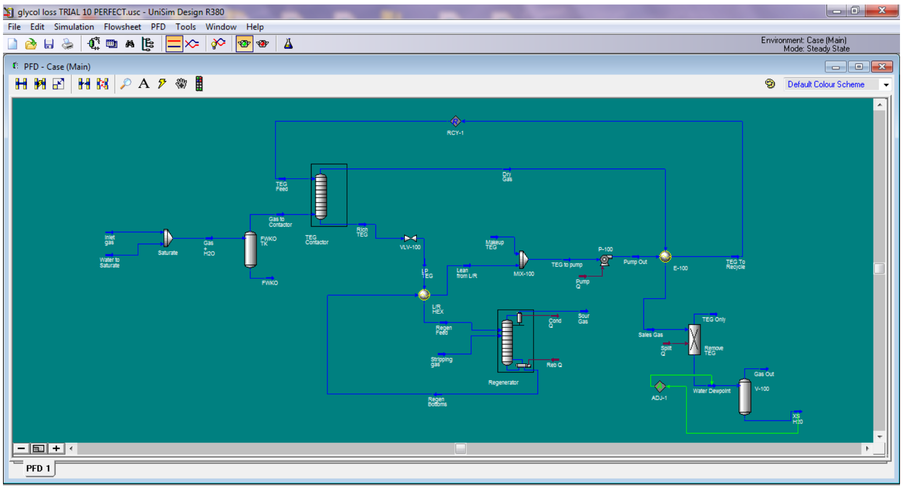
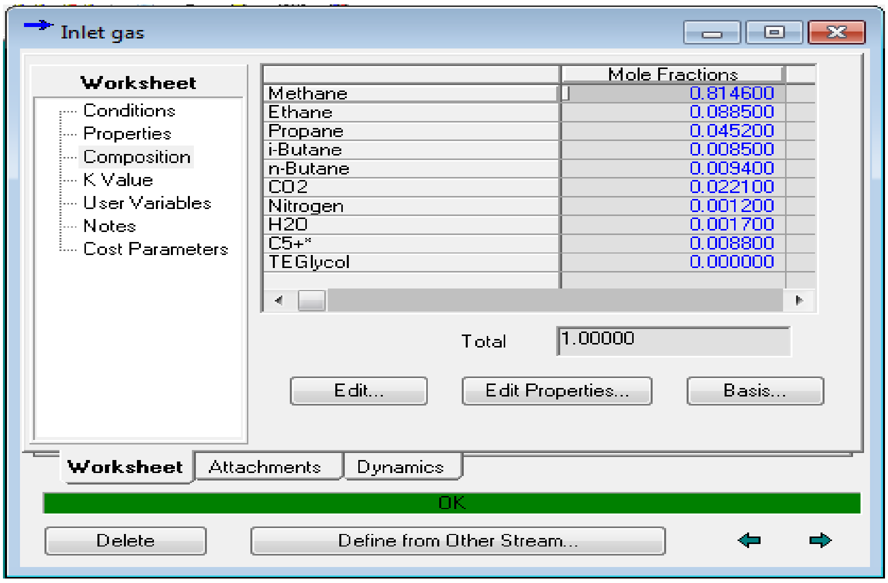


Figure 2.1 Process Flow Diagram of Hysys Simulation for a TEG dehydration process



	Mole Fractions
Methane	0.814600
Ethane	0.088500
Propane	0.045200
i-Butane	0.008500
n-Butane	0.009400
CO2	0.022100
Nitrogen	0.001200
H2O	0.001700
C5+*	0.008800
TEGlycol	0.000000

Total 1.00000

Figure 2.2 Screen shot of the Inlet Gas Composition

Table 3. Results of HYSYS simulation of maximum and minimum values for vaporization losses

S/N	GAS FLOW RATE (m3/hr)	GAS TEMP. (°C)	CONT PRESSURE (Bars)	LEAN GLYCOL CONC. (%)	GLYCOL CIRC. RATE (m3/hr)	LEAN GLYCOL TEMP. (°C)	REFLUX RATE (kgmole/hr)	REBOILER TEMP. (°C)	REBOILER PRESSURE (Bars)	GLYCOL LOSS (m3/hr)
128	400	55	83	99.7	25	60	0.2	204	0.6	-0.2457
129	400	55	83	99.7	25	60	0.4	204	0.6	-0.2479
130	400	55	83	99.7	25	60	0.6	204	0.6	-0.2492
131	400	55	83	99.7	25	60	0.8	204	0.6	-0.2516
132	400	55	83	99.7	25	60	1	204	0.6	-0.253
133	200	25	34	95	6	10	0.2	182	0.1	1.8392
134	200	25	34	95	6	10	0.4	182	0.1	1.8376
135	200	25	34	95	6	10	0.6	182	0.1	1.8378
136	200	25	34	95	6	10	0.8	182	0.1	1.8383
137	200	25	34	95	6	10	1	182	0.1	1.8392

IV. 12-RUN PLACKETT-BURMAN DESIGN OF EXPERIMENT

In order to carry out the Design of Plackett-Burman 12-run Experiment, eleven of the significant glycol dehydration process parameters were selected along with their corresponding high and low values. These were then used as input in the customized Plackett–Burman 12-Run matrix. Table 4 shows the key process variables that were engaged in the experiment.

Table 4. Vaporization Loss: Determinant Independent Process Variables

S/N		VARIABLES	UNITS	MIN	MAX
1	A	Gas Flow Rate	Std m^3/hr	200	400
2	B	Gas Temperature	^{O}C	25	40
3	C	Contactor Pressure	Bars	34	83
4	D	Lean Glycol Concentration	%	95	99,7
5	E	Lean Glycol Circulation Rate	m^3/hr	6	25
6	F	Lean Glycol Temperature	^{O}C	10	60
7	G	Condenser Temperature	^{O}C	60	100
8	H	Reboiler Pressure	Bars	0.1	0.6
9	I	Reboiler Temperature	^{O}C	182	204
10	J	Stripping Gas Rate	Std m^3/m^{3}TEG	0	90
11	K	Number of Trays	No.	4	12

The results of the HYSYS simulation for glycol losses using the maximum and minimum values (two-level factorial design) were then used as input for the 12- run Plackett-Burman Design of Experiment matrix for regression analysis. This yielded the predictive model for evaluating the in-process glycol losses by vaporization from both the contactor and regenerator. The equation is linear with a coefficient of determination $R^2 = 1$ (indicating a perfect fit), and because of this, the analysis of variance (ANOVA) yielded a zero residual. The zero-residual result is a pointer to the fact that all the selected factors are active to some degree in the process. None of them being inert. The results are presented in the tables 5 to 7.

Table 5. Table. Plackett-Burman 12-Run Output Matrix

S/N	**Q_G**	**T_G**	**$P_{Contac.}$**	**cTEG**	**Q_{TEG}**	**T_{TEG}**	**$T_{Cond.}$**	**P_R**	**T_R**	**S_{Gas}**	**N**	**Avg**	**Range**	**Response**
1	400	40	34	99.7	25	60	99	0.1	174	90	6	5.97	0.000	5.974
2	6	400	40	34	99.7	25	60	99	0.1	174	90	0.79	0.000	0.7891
3	90	6	400	40	34	99.7	25	60	99	0.1	174	0.00	0.000	0.0024
4	174	90	6	400	40	34	99.7	25	60	99	0.1	0.00	0.000	0.0022
5	0.1	174	90	6	400	40	34	99.7	25	60	99	-0.04	0.000	-0.039
6	99	0.1	174	90	6	400	40	34	99.7	25	60	5.99	0.000	5.992
7	60	99	0.1	174	90	6	400	40	34	99.7	25	0.85	0.000	0.8453
8	25	60	99	0.1	174	90	6	400	40	34	99.7	0.01	0.000	0.0056
9	99.7	25	60	99	0.1	174	90	6	400	40	34	16.48	0.000	-16.48
10	34	99.7	25	60	99	0.1	174	90	6	400	40	0.78	0.000	0.7797
11	40	34	99.7	25	60	99	0.1	174	90	6	400	1.48	0.000	1.484
12	200	25	34	95	6	10	99	0.1	174	0	6	0.02	0.000	0.0168

Table 6. Summary Output

SUMMARY OUTPUT		Force Constant to Zero
		FALSE
Regression Statistics		
Multiple R	1.000	
R Square	1.000	Goodness of Fit >= 0.80
Adjusted R Square	#DIV/0!	
Standard Error	0.000	
Observations	12	

Table 7. Summary of Regression Analysis

	Coefficients	***Standard Error***	***t Stat***	***P-value***	***Lower 95%***	***Upper 95%***	***Lower 99%***	***Upper 99%***
Intercept	7.600330557	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
Q_{Gas}	0.031786157	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
T_{Gas}	-0.008055869	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
$P_{Contac.}$	-0.010416164	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
cTEG	-0.018107393	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
Q_{TEG}	-0.00690994	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
T_{TEG}	0.014625505	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
$T_{Cond.}$	-0.001690905	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
P_R	-0.010817429	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
T_R	-0.066785503	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
Q_{strip}	-0.008823055	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
N	0.003225015	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
Average	0	0	0	#NUM!	#NUM!	#NUM!	#NUM!	#NUM!
Range	0	0	0	#NUM!	#NUM!	#NUM!	#NUM!	
Q_{gvl} = 7.6 +0.032* Q_{Gas} -0.008* T_{Gas} -0.01* P_{Contac} -0.018* cTEG -0.007* Q_{TEG} +0.015* T_{TEG} -0.002* $T_{Cond.}$ -0.011* P_R -0.067* T_R -0.009* Q_{strip} +0.003* N								

The model for the evaluation of glycol loss by vaporization is obtained as:

Q_{gvl} = 7.6 +0.031786157* Q_{Gsa} - 0.008055869* T_{Gas} -0.010416164* $P_{Contac.}$ -0.018107393* cTEG - 0.00690994* Q_{TEG} + 0.014625505* T_{TEG} -0.001690905* $T_{Cond.}$ -0.010817429* P_R -0.066785503* T_R - 0.008823055* Q_{strip}+0.003225015* N

Where,

Q_{gvl} = Glycol Loss by Vaporization
Q_{Gas} = Gas Flow Rate
T_{Gas} = Gas Temperature
P_{Contac} . = Contactor Pressure
cTEG = Lean Glycol Concentration
Q_{TEG} = Lean Glycol Circulation Rate
T_{TEG} = Lean Glycol Temperature
T_{Cond} = Condenser Temperature
P_R = Reboiler Pressure
T_R = Reboiler Temperature
Q_{strip} = Stripping Gas Rate
N = Number of Trays

Validation of Model

A comparison of measured value with the calculated values showed an excellent match, validating the coefficient of determination R^2 =1. Table 8

Table 8. Measured (M value) compared with calculated (Cal value)

A	**B**	**C**	**D**	**E**	**F**	**G**	**H**	**I**	**J**	**K**	**M value**	**Cal value**
450	40	34	99.7	25	60	99	0.1	174	90	6	7.5633	7.5633
425	40	34	99.7	25	60	99	0.1	174	90	6	6.7687	6.7687
400	40	34	99.7	25	60	99	0.1	174	90	6	5.974	5.974
390	40	34	99.7	25	60	99	0.1	174	90	6	5.6561	5.6561
380	40	34	99.7	25	60	99	0.1	174	90	6	5.3383	5.3383
370	40	34	99.7	25	60	99	0.1	174	90	6	5.0204	5.0204
350	40	34	99.7	25	60	99	0.1	174	90	6	4.3847	4.3847
300	40	34	99.7	25	60	99	0.1	174	90	6	2.7954	2.7954
250	40	34	99.7	25	60	99	0.1	174	90	6	1.2061	1.2061
200	40	34	99.7	25	60	99	0.1	174	90	6	-0.3832	-0.3832

180	40	34	99.7	25	60	99	0.1	174	90	6	-1.0190	-1.0190
0.0318	-0.0081	-0.0104	-0.0181	-0.0069	0.0146	-0.0017	-0.0108	-0.0668	-0.0088	0.0032		
Intercept	7.600330557											

V. CONCLUSION

Measuring glycol losses can be quite challenging and oftentimes imprecise. However, a working knowledge of what can be expected will go a long way in mitigating these losses. What cannot be measured cannot be managed. The major loss of glycol from a dehydration plant is due to mechanical and carryover losses from the absorption column, the flash tank or the regeneration column. These can be measured by sampling downstream of the vessels. In this research, an appropriate predictive model equation that offers support for oil and gas engineers towards a fast computation of TEG vaporization losses is presented. It is vital to note that the systematic analytical approaches offered by Hysys and Plackett-Burman Design of Experiments resulted in a smooth equation. But optimum conditions and results generated from sensitivity analysis in most cases cannot be directly applied to real plant operations. Oftentimes, a modification in plant design and diligent unit operations are essential in minimizing glycol losses while maintaining the required moisture content as specified by the gas transmission company.

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