A PERFORMANCE STUDY FOR WESTERN DESERT GAS COMPLEX (WDGC) UNDER VARIOUS FEED CONDITIONS

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ABSTRACT

The train-C at Western Desert Gas Complex (WDGC) is NGLs (Natural Gas Liquids) recovery train located in Alexandria - Egypt. The train-C employs a mix of Refrigeration and Turbo-expander technologies as a Gas Sub-cooled Process (GSP) to NGL/LPG recoveries with fractionating the raw gas into residue gas, ethane/Propane Mixture, Propane, LPG, and Condensates products based upon boiling point differences. The train-C handles 300 MMSCFD and designed to maximize the recovery of produced Ethane and Propane mixture.

The plant's feed Gas composition & conditions at train-C is changing over time which leads to make changes in plant operating conditions. Feed gas pressure decreased to 66 Kg/cm^2 compared with 70.5 Kg/cm^2 in design , an increase in feed CO_2 content to be 4.8% instead of 3.7% and also an increase in feed methane (C_1) content to be 82% instead of 79% which caused decrease Ethane/ Propane recoveries in current actual operating conditions.

In this paper, the train-C is simulated using a steady state simulator (Aspen HYSYS) & Peng–Robinson (PR) equation of state (EoS) considering Equipment's design data and products specifications. The simulation was made with a number of possibilities options and optimizations can be done to produce additional C_2 with/without simple retrofitting plant's equipment's & facilities. Heat integration with narrow minimum temperature approach for heat exchangers, reducing De-methanizer column pressure and change split going to the sub-cooler exchanger / turbo-expander are studied, few enhancement to ethane recovery is occurred. With installing new vessel to separate GSP stream leaded to decreasing vapor load in the upper section of De-methanizer which increased ethane recovery to acceptable level.

Keywords: - NGLs, GSP, Ethane recovery, Refrigeration technology, Turbo-expander technology

INTRODUCTION

GASCO established the train-C at WDGC to maximize the recovery of the ethane and propane from the existing western desert (W/D) gas fields located at Alexandria, Egypt.

The complex produces C_2/C_3 gas mixture, which is utilized as a feedstock to the ethylene production plant (SIDPEC), commercial propane, low/high vapor pressure LPG, sales gas and condensate.

Train-C is designed to handle a total feed capacity of 300 MMSCFD and comprising the following:

- Dehydration package is designed to handle feed gas with 150 ppm inlet water content to satisfy the water dew point requirement.
- Chilling train & Conventional propane refrigeration package
- Turbo-expander system comprising 2 * 50% identical trains.
- Demethanizer column.
- Fractionation train comprising deethanizer, depropanizer, debutanizer columns and associated equipments (overhead condensers, overhead accumulators, bottom pumps, fired heaters,..etc)
- Sales gas compression train.

The plant performance is guaranteed to recover 77% of ethane, 95% of propane, and 95% of butane in feed gas and controlled by the feed gas composition and conditions or flow rate of supplied to the plant during the performance test run. In addition, guarantee covers products specifications as will be illustrated later.

The paper will include the following items:-

- Design operating conditions of Train-C as a base case for the plant performance.
- Current train-C operating conditions of two cases; high CO_2 content feed gas & high Methane content feed gas.
- Plant simulation runs using HYSYS process simulation software version 7.2 for two cases of current feed gas conditions & without any retrofitting in plant by;
    - ➢ Lowering operating pressure at demethanizer column

- Narrow minimum temperature approach for heat exchangers taking into account the CO_2 freeze-out at the top of demethanizer

- Changing the gas split from low temperature separator between the sub-cooler exchanger and turbo-expander fixed with design capacities of both equipments.

- Train C simulation run with current two cases feed gas conditions and installing new vessel at downstream of demethanizer condenser to separate lean gases from gas sub-cooled stream, which lead to decrease the vapor load in the upper section of the demethanizer column (10-C-02).

- Analysis of the results of different simulation runs cases.

PROCESS DESCRIPTION

Chilling section

The feed gas stream coming from the molecular sieve dehydration package is directed to the demethanizer reboiler (10-R-03) where is cooled using the demethanizer bottom tray (tray-38) liquids.

Cold feed outlet from 10-R-03 is further cold by heat exchanger heat in the gas/gas heat exchanger (10-E-03 A/B) using demethanizer OVHD gas stream coming from 10-E-04 A/B. the precooled feed gas out from 10-E-03 A/B is further cooled in the propane chiller (12-E-01) which is a part of the new propane refrigeration package (10-X-02).

The propane refrigeration package (10-X-02) is necessary to provide refrigeration for the process stream outlet from the gas/gas exchanger (10-e-03 A/B) via propane chiller (12-E-01) and to the deethanizer OVHD gas via deethanizer OVHD condenser (12-E-04).

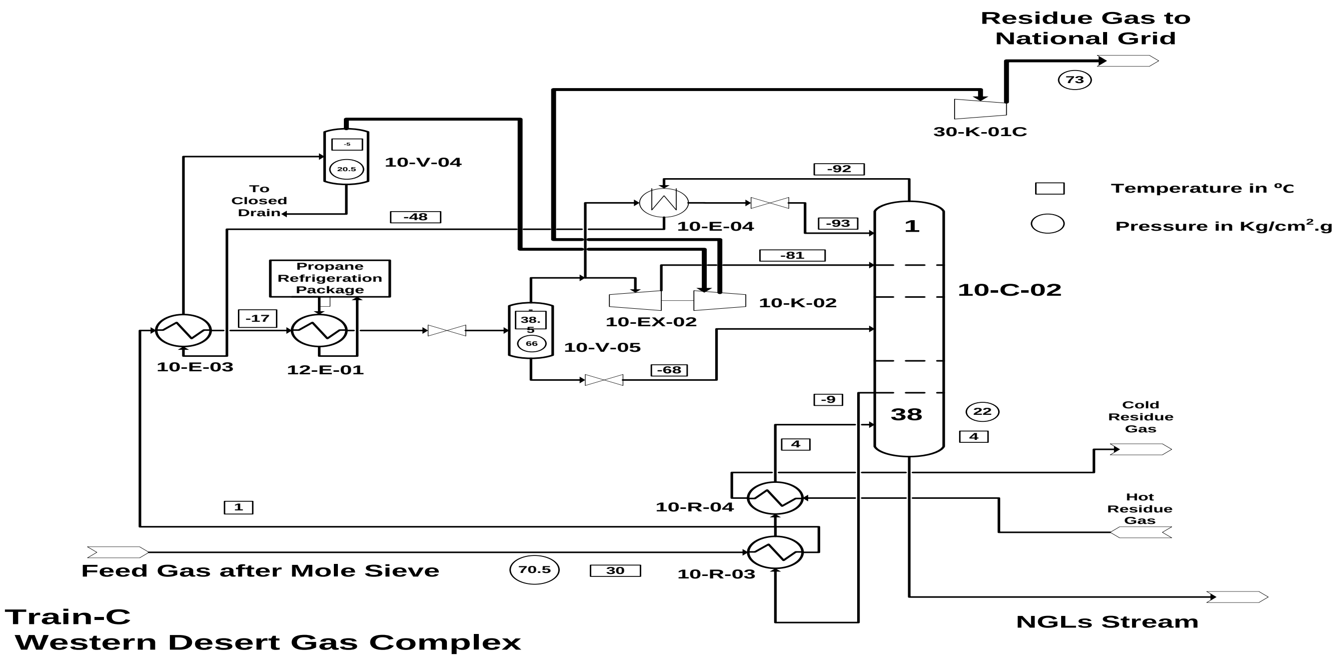
Refrigerated feed gas outlet from (12-E-01) is routed under flow control valve to the low temperature separator (10-V-05) where liquids are separated and directed under level resetting flow control to the demethanizer tower (10-C-02) at tray-24.

Flashed vapors from (10-V-05) are splitted into two streams. The first is cooled by cross exchange with the demethanizer OVHD vapors in the demethanizer condenser (10-E-04 A/B) and then it is directed through a flow control valve to the demethanizer tower (10-C-02) at the top tray where vapor gas straight overhead and the liquid forms the reflux to tray 1 of the top section of the demethanizer (10-C-02). The second stream is directed to two identical parallel turbo-expander trains (10-EX-02 A/B) and then to the demethanizer tower (10-C-02) at tray-13.

Recovery section

The demethanizer tower (10-C-02) is a trayed tower with 38 trays which separate methane (C_1) from ethane and heavier compounds (C_2^+). The tower is operated at pressure of (19.6-20.5) Kg/cm^2.g approximately. Liquid from tray-38 (bottom tray) is heated for demethanizer reboiling purpose by cross exchange with dehydrated feed gas in the demethainzer reboiler (10-R-03)., a trim reboiler (10-R-04) working in series with the main reboiler (10-R-03) is provided. Heating medium will be the hot residue gas taken and returned.

OVHD vapors discharged from (10-E-04 A/B) tube side are directed to the gas/gas exchanger (10-E-03 A/B). Warm up gas outlet from the gas/gas exchanger (10-E-03 A/B) are directed to the re-compressor Knock-Out Drum (10-V-04) to separate any entrained liquid in the gas stream (if any) and then compressed to (20.8-21.8) Kg/cm2.g in the expander re-compressor (10-K-02 A/B). The expander (10-EX-02 A/B) drives the re-compressor (10-K-02 A/B).



PFD of Train-C at WDGC

DESIGN OPERATING CONDITIONS OF THE PLANT (BASE CASE)

Key streams; as per the plant design operating conditions, the plant mainstreams as following:

Conditions & Composition (Mole %)	Feed Gas	Residue Gas	NGLs Stream
Vapour Fraction	1.00	1.00	0.00
Temperature [C]	30	3	-4
Pressure [kg/cm^2_g]	70.5	22.8	17.8
Molar Flow [MMSCFD]	300	249	51
Mass Flow [tonne/d]	7419	5058	2360
Liquid Volume Flow [barrel/day]	132989	101240	31754
H_2O	0.00%	0.00%	0.00%
Nitrogen	0.81%	0.97%	0.00%
CO_2	3.70%	1.89%	12.61%
H_2S	0.00%	0.00%	0.00%
Methane	79.22%	95.23%	0.30%
Ethane	9.90%	1.83%	49.71%
Propane	4.45%	0.08%	26.01%
i-Butane	0.71%	0.00%	4.18%
n-Butane	0.91%	0.00%	5.39%
i-Pentane	0.10%	0.00%	0.60%
n-Pentane	0.10%	0.00%	0.60%
n-Hexane	0.04%	0.00%	0.24%
n-Heptane	0.03%	0.00%	0.18%
n-Octane	0.03%	0.00%	0.18%

Hydrocarbon recovery percent in demethanizer bottom stream:

Ethane	84.66%	**i-Butane**	99.27%
Propane	98.55%	**n-Butane**	99.87%

Design operating conditions of main equipments of NGLs recovery section

Unit Operation		Input Stream	Temp. [C]	Pres.[kg/cm^2_g]	Flow Rate [kg/h]	Output Stream	Temp. [C]	Pres. [kg/cm^2_g]	Flow Rate [kg/h]
Demethanizer Reboiler (10-R-03);		**Feed Gas**	30	70.5	309119.3	**400 A**	1	70.33	309119.3
Design LMTD, C	**14**								
Design Duty, MM Kcal/hr (17.85% overdesign)	**6.68**								
Demethanizer Trim Reboiler (10-R-04);		**Liquid from De-C1 bottom**	2	22.5	173625.9	**Hot liquid stream**	4	22	173625.9
Design LMTD, C	**51**								
Design Duty, MM Kcal/hr, (14.62% overdesign)	**4.55**								
Gas/Gas Exchanger (10-E-03A/B);		**400 A**	1	70	309119.3	**400 B**	-17	69.89	309119.3
Design LMTD, C	**15.2**								
Design Duty, MM Kcal/hr (17.31% overdesign)	**6.75**								
Propane Chiller (12-E-01);		**400 B**	-17	69.5	309119.3	**400 C**	-37	69.06	309119.3
Design LMTD, C	**8.58**								
Design Duty, MM Kcal/hr (15% overdesign)	**8.63**								
Low Temperature Separator - LTS (10-V-05)		**400 D**	-38.6	66	309119.3	**400 E**	-38.6	66	218055.6
						407 A	-38.6	66	91063.7

Pressure,kg/cm2_g	**66**								
Temperature, C	**-38.5**								
Turbo-Expander (10-EX-02A/B)& Expander Recompressor (10-K-02A/B);		**401**	-38.6	66	65416.7	**402**	-81	21.8	65416.7
Design Flow rate , Kg/hr	**65900**								
Design Discharge Pressure, kg/cm2_g	**20.8**								
Demethanizer Condenser (10-E-04 A/B);		**400 F**	-38.6	66	152638.9	**403 A**	-57.5	65.89	152638.9
Design LMTD, C	**18.5**								
Design Duty, MM Kcal/hr, (25.65% overdesign)	**6.079**								
Demethanizer Column (10-C-02);		**402**	-81	21.8	65416.7	**404 A**	-92	21.8	210770.6
		403	-93	21.6	152638.9	**410 A**	4	22	98348.8
		407	-68	22.1	91063.7				
Bottom Operating pressure, kg/cm2_g	**22**								
Tray-1 Temperature, C	**-92**								
Tray-38 Temperature, C	**4**								

CURRENT KEY PARAMETERS OF TRAIN-C OPERATING CONDITIONS COMPARED WITH DESIGN

Feed Gas Conditions & Composition

Conditions & Composition (Mole%)	Design	Current Case-1:high CO_2 %	Current Case-2: high C_1%
Vapour Fraction	1	1	1
Temperature [C]	30	20.58	25
Pressure [kg/cm^2_g]	**70.5**	**65.88**	**66**
Molar Flow [MMSCFD]	300	300	300
Mass Flow [tonne/d]	7419	7165	7196
H_2O	0.00%	0.00%	0.00%
Nitrogen	0.81%	0.64%	0.63%
CO_2	**3.70%**	**4.69%**	**4.03%**
H_2S	0.00%	0.00%	0.00%
Methane	**79.22%**	**81.06%**	**82.06%**
Ethane	9.90%	8.57%	8.51%
Propane	4.45%	3.41%	3.21%
i-Butane	0.71%	0.53%	0.51%
n-Butane	0.91%	0.75%	0.70%
i-Pentane	0.10%	0.15%	0.14%
n-Pentane	0.10%	0.11%	0.11%
n-Hexane	0.04%	0.04%	0.03%
n-Heptane	0.03%	0.03%	0.03%
n-Octane	0.03%	0.02%	0.03%

VARIATION BETWEEN CURRENT & DESIGN OPERATING CONDITION

- Currently feed gas pressure is around 66 compared with 70.5 Kg/cm^2.g as a design feed gas pressure, CO_2 in actual feed gas between 4 & 4.7% while in design 3.7 and also C_1 is 81 to 82.1% compared with 79.2% in design.

- Propane refrigeration package is running with 66% of its capacity, due to one-of-three Propane refrigerant compressor did not start-up yet which lead to chilling through main propane chiller around 66% of the required cooling duty.

THE IMPACT OF VARIATION IN TRAIN-C OPERATING CONDITION

Comparison Criteria	Current Case-1 (high CO_2)	Current Case-2 (high C_1)	Design	Remarks
Feed Gas Temperature, C	20.6	25	30	-Design for summer lean case.
Feed Gas Temp after Demethanizer reboiler, C	4	4	1	
Temperature of Inlet Stream to Propane Chiller (after Gas/Gas Exchanger), C	-20	-20	-17	
Propane Chiller duty, MM Kca/hr	5.5	5.3	7.5	- Due to shortage in Refrigerant compression power
Temperature of Outlet Stream to Propane Chiller, C	-36	-36	-37	
Low Temp Separator Pressure, $Kg/cm^2.g$	58.5	58.5	66	
Low Temp Separator Temp, C	-39.5	-39.6	-38.6	
Vapor stream from LTS, Kg/hr	245004	247368	218055.6	Maximum Capacity 243685 Kg/hr
ΔP through Turbo-Expander, Kg/cm^2	33.5	33.5	44	- Due to lowering in feed gas pressure
Expander downstream pressure, $Kg/cm^2.g$	25	25	21.8	
Expander downstream Temperature, C	-72.2	-72.4	-81	
Liq stream from LTS to Demethanizer Temp, C	-60.7	-60.6	-68	
Gas Subcooled Stream Temp (Oulet of Demethanizer Condneser & Inlet to Demethanizer), C	-53 & -78.4	-53 & -78.4	-57.5 & -93	
Demethanizer Top Temp, C	-77.25	-77.3	-92	
Demethanizer Top Pressure, $Kg/cm^2.g$	24.8	24.8	21.5	
Demethanizer Bottom Temp, C	16	17.7	4	
Demethanizer Bottom Pressure, $Kg/cm^2.g$	25.22	25.22	22	
C2 Recovery %	59.6	57.8	84.66	
C3 Recovery %	93.7	93.2	98.55	

SIMULATION MODELS OF THE POSSIBILITIES OPTIONS AND OPTIMIZATIONS CAN BE DONE TO INCREASE C_2&C_3 RECOVERIES

Simulation models were carried out using the HYSYS simulator and the PR (Peng–Robinson) equation of state for the calculation of thermodynamic properties.

Basis of Simulation runs:-

- All Basis of Design should be applied.

- All equipment's operating conditions will be within Design specifications.

The possibilities options to increase C_2&C_3 recoveries can be done through the following four simulation models:

Case-1 (High CO_2 %):

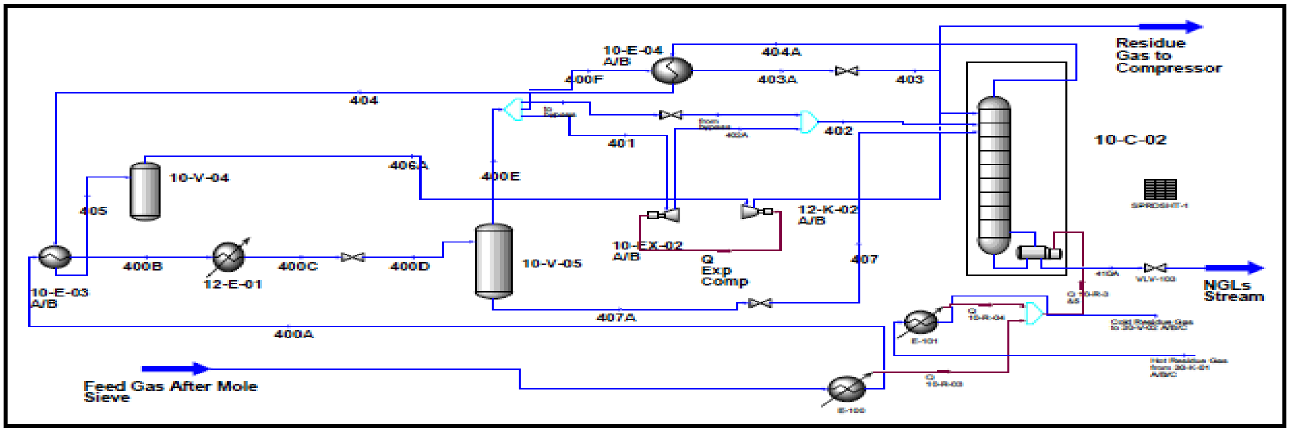
Case-1A: Lowering operating pressure at demethanizer column , narrow minimum temperature approach for heat exchangers taking into account the CO_2 freeze-out at the top of demethanizer and manipulate the gas split from low temperature separator between the sub-cooler exchanger and turbo-expander fixed with design capacities of both equipments (without any retrofitting in existing equipments).

Case-1B:. with retrofitting existing facilities by installing new vessel at downstream of demethanizer condenser to separate lean gases from gas sub-cooled stream, which lead to decrease the vapor load in the upper section of the demethanizer column (10-C-02).

Case-2 (High C_1 %):

Case-2A: Lowering operating pressure at demethanizer column , narrow minimum temperature approach for heat exchangers taking into account the CO_2 freeze-out at the top of demethanizer and manipulate the gas split from low temperature separator between the sub-cooler exchanger and turbo-expander fixed with design capacities of both equipments (without any retrofitting in existing equipments).

Case-2B:. with retrofitting existing facilities by installing new vessel at downstream of demethanizer condenser to separate lean gases from gas sub-cooled stream, which lead to decrease the vapor load in the upper section of the demethanizer column (10-C-02).



PFD of Cases 1A & 2A

Case-1 (High CO_2 % in feed gas):

A- Lowering operating pressure at Demethanizer Column , narrow minimum temperature approach (LMTD) for heat exchangers and manipulate the gas split from low temperature separator between demethanizer condenser

With applying the following operating conditions;

Heat Exchanger	Current Duty & LMTD, M Kcal/hr & C	Case-1-A Duty & LMTD, M Kcal/hr & C	Design Reference
Demethanizer Reboiler 10-R-03	3.21 & 16	3 & 15	Normal duty 5.67 & Design 6.68 LMTD Min 14
Gas/Gas Exchanger 10-E-03 A/B	5.9 & 20	6.2 & 17	Normal duty 5.75 & Design 6.75 LMTD Min 15.2
Demethanizer Condneser 10-E-04 A/B	3.2 & 19	3.74 & 20 (To obtain LMTD for other exchangers)	Normal duty 4.84 & Design 6.1 LMTD Min 18.5

Demethanizer Column	Current Case	Case-1-A	Design Reference
Operating Pressure at top & bottom, $Kg/cm^2.g$	24.8 & 25.2	22 & 22.5	Normal 22 (Design Pressure 35)

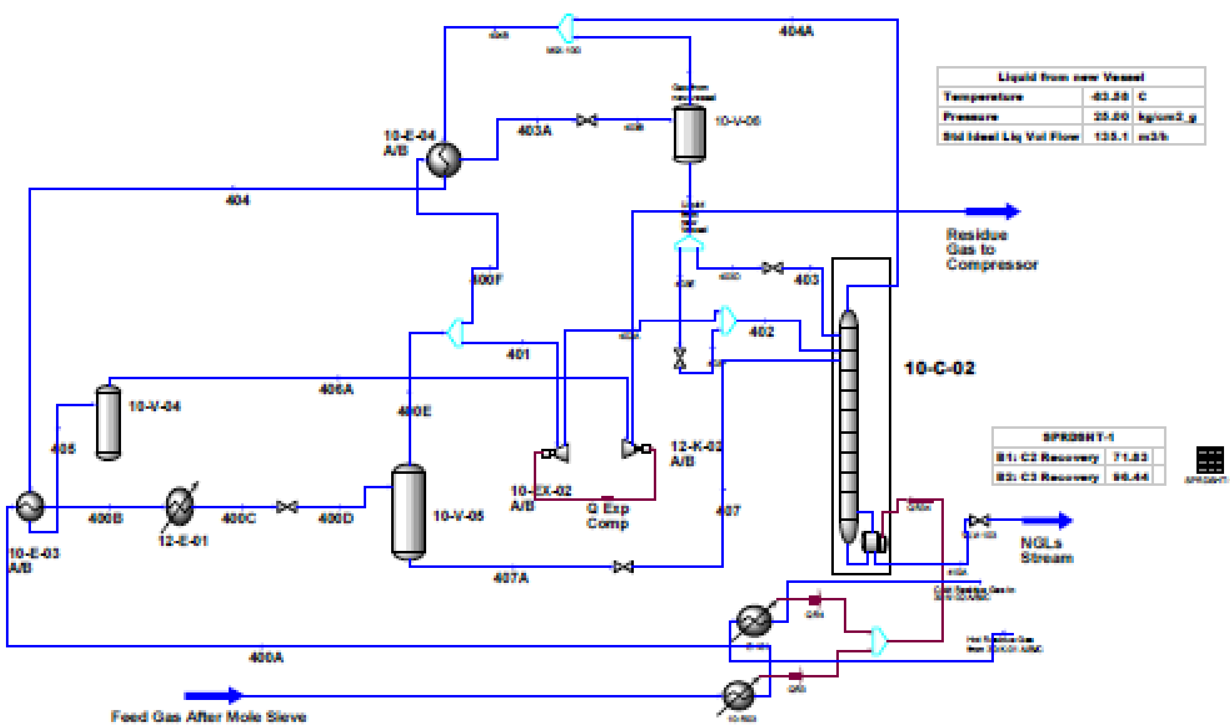
Equipment	Current Case	Case-1-A	Design Reference
Gas from LTS to Demethanizer Condenser 10-E-04 A/B, Kg/hr	167900	166418.4	Normal 152638.9 Maximum 167900
Gas from LTS to Turbo-Expander (10-EX-02 A/B) & power, Kg/hr & KW	75785 & 875	75785 & 970.6	Normal 65416.7 & 845 Maximum 75785 & 972
Gas from LTS to By-Pass Valve (212-PV-223A/B), Kg/hr	1319	0	0 in Normal

- **In case-1-A; Gas stream from LTS is 242203.4 Kg/hr compared with 245004 Kg/hr in current case-1, therefore no need to bypass any quantities.**

Case-1-B: installing new vessel at downstream of demethanizer condenser, manipulate the gas split from low temperature separator between demethanizer condenser, and exchanger and turbo-expander Lowering operating pressure at Demethanizer Column and narrow minimum temperature approach (LMTD) for heat exchangers

With installing new vessel (10-V-06) downstream of shell stream of Demethanizer condenser (12-E-04 A/B), applying the following operating conditions to achieve the maximum available enhancement:-

- Outlet temperature of process stream from Propane main chiller (12-E-01) is -37^{0}C.
- Pressure difference between the new vessel (10-V-06) and the demethanizer column (10-C-02) is 3 bar.
- Liquid stream from new vessel (10-V-06) is splitting to two parts; one part (around 100 m^3/hr) to top tray of demethanizer column (10-C-02) and 2nd part mix with expander outlet stream entering the demethanizer.



PFD of Cases 1B & 2B

Case-2 (High C_1 % in feed gas):

A- Lowering operating pressure at Demethanizer Column , narrow minimum temperature approach (LMTD) for heat exchangers and manipulate the gas split from low temperature separator between demethanizer condenser

With applying the following operating conditions;

Heat Exchanger	Current Duty & LMTD, M Kcal/hr & C	Case-2-A Duty & LMTD, M Kcal/hr & C	Design Reference
Demethanizer Reboiler 10-R-03	4 & 16	3.8 & 15	Normal duty 5.67 & Design 6.68 LMTD Min 14
Gas/Gas Exchanger 10-E-03 A/B	5.7 & 22	6.33 & 20	Normal duty 5.75 & Design 6.75 LMTD Min 15.2
Demethanizer Condneser 10-E-04 A/B	3.12 & 19	3.66 & 20 (To obtain LMTD for other exchangers)	Normal duty 4.84 & Design 6.1 LMTD Min 18.5

Demethanizer Column	Current Case	Case-2-A	Design Reference
Operating Pressure at top & bottom, $Kg/cm^2.g$	24.8 & 25.2	22 & 22.5	Normal 22 (Design Pressure 35)

Equipment	Current Case	Case-2-A	Design Reference
Gas from LTS to Demethanizer Condenser 10-E-04 A/B, Kg/hr	167900	167900 (High)	Normal 152638.9 Maximum 167900
Gas from LTS to Turbo-Expander (10-EX-02 A/B) & power, Kg/hr & KW	75785 & 875	75785 & 972	Normal 65416.7 & 845 Maximum 75785 & 972
Gas from LTS to By-Pass Valve (212-PV-223A/B), Kg/hr	3683	1027	0 in Normal

- **In case-2-A; Gas stream from LTS is 244712 Kg/hr compared with 247368 Kg/hr in current case-2, therefore no need to bypass any quantities.**

Simulation Results; Case-1 (High CO_2 % in feed gas):

Comparison Criteria	**Current Case-1**	**Case-1 A**	**Case-1 B Adding new Vessel**	**Design**
Feed Gas Flow	300	300	300	300
Feed Gas Pressure, $Kg/cm^2.g$	65.88	65.88	65.88	70.5
Feed Gas Temperature, C	20.6	20.6	20.6	30
Feed Gas Temp after Demethanizer reboiler, ^0C	4	5	1	1
Temperature of Inlet Stream to Propane Chiller (after Gas/Gas Exchanger), ^0C	-20	-20.5	-19.8	-17
Propane Chiller duty, MM Kca/hr	5.5	5.7	5.9	7.5
Temperature of Outlet Stream to Propane Chiller, ^0C	-36	-37	-37	-37.4
Low Temp Separator Pressure, $Kg/cm^2.g$	58.5	59.2	62.5	66
Low Temp Separator Temp, ^0C	-39.5	-40	-38.23	-38.6
Vapor stream from LTS, Kg/hr	245004	242203.4	245113	Normal 218055.6 **Maximum 243685**
ΔP through Turbo-Expander, Kg/cm^2	33.5	36.5	39.34	44
Expander downstream pressure, $Kg/cm^2.g$	25	22.7	23.15	21.8
Expander downstream Temperature, ^0C	-72.2	-76.2	-75.73	-81
Liq stream from LTS to Demethanizer Temp, C	-60.7	-65.1	-64.85	-68
Gas Subcooled Stream Temp (Oulet of Demethanizer Condneser & Inlet to Demethanizer), ^0C	-53 & -78.4	-55 & -84	-55 & -87	-57.5 & -93
Demethanizer Top Temp, ^0C	-77.25	-83	-84	-92
Demethanizer Top Pressure, $Kg/cm^2.g$	24.8	22	21.83	21.5
CO_2 freeze-out temp at top of Demethanizer , ^0C	-84.2	-85.7	-85.45	-91
Demethanizer Bottom Temp, ^0C	16	7.5	4.7	4
Demethanizer Bottom Pressure, $Kg/cm^2.g$	25.22	22.5	22.12	22
C_2 Recovery %	**59.6**	**67.2**	**71.83**	**84.66**
C_3 Recovery %	**93.6**	**95.9**	**96.44**	**98.55**
NGLs Stream , Ton/day	**1632**	**1764**	**1852**	**2360**

C_2 Flow, Ton/day	**550.5**	**622**	**664**	**903.8**
C_3 Flow, Ton/day	**505**	**517**	**520**	**693.5**
Yield NGLs compared with current flow, Ton/day	**-**	**132**	**220**	**728**
Yield NGLs of installing new vessel, Ton/day	**-**	**-**	**88**	**-**

Simulation Results; Case-2 (High C_1 % in feed gas):

Comparison Criteria	Current Case-2	Case-2 A	Case-2 B Adding new Vessel	Design
Feed Gas Flow	300	300	300	300
Feed Gas Pressure, Kg/cm^2.g	66	66	66	70.5
Feed Gas Temperature, C	25	25	25	30
Feed Gas Temp after Demethanizer reboiler, ^{0}C	4	5	1	1
Temperature of Inlet Stream to Propane Chiller (after Gas/Gas Exchanger), ^{0}C	-20	-21.5	-22.2	-17
Propane Chiller duty, MM Kca/hr	5.3	5.3	5	7.5
Temperature of Outlet Stream to Propane Chiller, ^{0}C	-36	-37	-37	-37.4
Low Temp Separator Pressure, Kg/cm^2.g	58.5	59	58	66
Low Temp Separator Temp, ^{0}C	-39.6	-40.3	-40.85	-38.6
Vapor stream from LTS, Kg/hr	247368	244712	243901	Normal 218055.6 **Maximum 243685**
ΔP through Turbo-Expander, Kg/cm^2	33.5	36.2	35.7	44
Expander downstream pressure, Kg/cm^2.g	25	22.83	22.3	21.8
Expander downstream Temperature, ^{0}C	-72.4	-76.22	-77	-81
Liq stream from LTS to Demethanizer Temp, C	-60.6	-65	-67	-68
Gas Subcooled Stream Temp (Oulet of Demethanizer Condneser & Inlet to Demethanizer), ^{0}C	-53 & -78.4	-55 & -84	-57.5 & -89	-57.5 & -93
Demethanizer Top Temp, ^{0}C	-77.3	-83	-85.7	-92
Demethanizer Top Pressure, Kg/cm^2.g	24.8	22	20	21.5
CO_2 freeze-out temp at top of Demethanizer , ^{0}C	-85.8	-87	-87	-91
Demethanizer Bottom Temp, ^{0}C	17.7	9.3	3.2	4
Demethanizer Bottom Pressure, Kg/cm^2.g	25.22	22.5	20.5	22
C_2 Recovery %	**57.8**	**65.7**	**70.3**	**84.66**

C_3 Recovery %	**93.2**	**95.5**	**96.6**	**98.55**
NGLs Stream , Ton/day	**1522.3**	**1644**	**1721**	**2360**
C_2 Flow, Ton/day	**531**	**603**	**646**	**903.8**
C_3 Flow, Ton/day	**473**	**485**	**490**	**693.5**
Yield NGLs compared with current flow, Ton/day	**-**	**122**	**199**	**838**
Yield NGLs of installing new vessel, Ton/day	**-**	**-**	**77**	**-**

CONCLUSION

- The current more lean feed gas leads to expander is bypassing certain gas flow rate and increase temperature of inlet streams to demethanizer, therefore Ethane & Propane recoveries decreased.

- The current vapor flow from LTS is high (247368 Kg/hr for high methane feed gas & 245004 for high CO_2 feed gas) compared with normal design flow 218056 Kg/hr or maximum design 243685 Kg/hr, due to more lean feed gas & shortage of propane chiller cooling.

- The current differential pressure through turbo-expander is low (33.5 Kg/cm^2) compared with normal design (39.3 Kg/cm^2), due to low feed gas pressure (66 Kg/cm^2) compared with normal design (70.5 Kg/cm^2).

- The Ethane recovery is lower than design case, due to the higher vapor flow from LTS and expander-bypass gas.

- Narrow minimum temperature approach at Demethanizer reboiler (10-R-03) & Gas/Gas Exchanger (10-E-03 A/B) and Lowering operating pressure at demethanizer column will increase Ethane plus recovery, which obtains around 120 Ton/day C_2^+ products for high methane content feed gas & 130 Ton/day C_2^+ products for high CO2 content feed gas, without any retrofitting in existing facilities.

- Installing new vessel to separate vapors at downstream demethanizer condenser will achieve around 199 & 220 Ton/day C_2^+ products for feed gas two cases high methane & high CO_2 respectively, compared with current operating conditions.

- Installing new vessel to separate vapors at downstream demethanizer condenser will achieve around 77 & 88 Ton/day C_2^+ products for feed gas two cases high methane & high CO_2 respectively, compared with current optimized operating conditions.

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