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Theoretical Sizing & Design of The Equipment of a 40 MMSCFD Natural Gas Processing Plant based on the operating condition of Titas Gas Field Location #A

Mohammed Rashnur Rahman, N.M. Aftabul Alam Bhuiya, and Md. Rasel Miah

Petroleum & Mining Engineering,
Shahjalal University of Science & Technology,
Sylhet-3114, Bangladesh

ABSTRACT: Natural gas processing is a complex industrial process designed to reduce impurities from raw natural gas by separation process to produce gas which is known as pipeline quality dry natural gas. In Titas Gas Field (Location #A), natural gas process plant is mainly designed to separate the water from the raw gas and make the gas transmittable to the consumer. Natural gas dehydration for Titas Gas Field mainly includes the separation of water from gas by Glycol Dehydration Process. Besides, a little amount of condensate is also separated during the dehydration process. Condensate of Titas Gas Field mainly contents hydrocarbon of C_3-C_4. As the absorbent, TEG (Tri-Ethylene Glycol) is the most preferable to use. In this paper, we represent the design of the equipment using in the glycol dehydration plant of Titas Gas Field, Location #A. Equipment sizing means calculating optimum height, diameter, number of tray, capacity, circulation rate etc. of each equipment. Calculation of height, diameter, number of tray capacity, circulating rate etc. of the equipment by using various tables, figures, charts, methods. In this paper we designed Inlet 3-phase separator, Glycol contactor tower, Lean-Rich glycol heat exchanger, Glycol circulation pump, Glycol flash separator (3-phase), Glycol regenerator (Re-boiler), Stripping still. It has to be mentioned that this is a theoretical design of the equipment for a glycol dehydration plant based on the operating condition of Titas Gas Field, Location #A which can be more efficient than the operating equipment.

KEYWORDS: Dehydration, Impurities, Absorbent, TEG, Separator and Operating Condition.

1 INTRODUCTION

This project study refers to size and design the equipment of a glycol dehydration plant based on the operating condition of Titas Gas Field (Location #A). This project work is mainly to establish a general way to design any glycol dehydration plant on several operating conditions based on the theory. This work will be a help to any further glycol dehydration plant design of several gas fields. Operating parameters are mainly the pressure, temperature, gas flow rate, water content in raw gas, desired amount of water content in processed gas, required amount of glycol to remove 1 lb water etc. There are total 18 gas wells in Titas Gas Field. Titas Gas Field (Location #A) is situated in Brahmanbaria includes well no-1, 3, 4, 5, 7. Well no-3 is known as dead well now, as it has stopped producing gas. At present, gas of well no-1, 4, 5, and 7 is processing in the glycol dehydration plant of Titas Gas Field (Location #A). We will consider a single well for simple calculation with a capacity of 40 MMSCFD. year.

2 OPERATING CONDITION OF TITAS GAS FIELD LOCATION #A

It is desired to process 40 MMSCFD natural gas (mainly CH_4) containing water and hydrocarbon condensate. To remove 1lb water content from the raw gas, 3-4 gallon TEG is needed for the dehydration process.
The conditions for the inlet and outlet of the process plant are described here.
Feed gas conditions for the plant:
Inlet pressure: 1400-1500 psi

Inlet temperature: 140-150^{0}F
Flow rate: 35-40 MMSCFD
Water content in raw natural gas: 0.85 bbl/MMSCF
Condensate content: 0.80 bbl/MMSCF

The Dehydration plant should process the natural gas according to the following basis:
Outlet pressure: 1000 psi
Outlet temperature: 120^{0}F
Maximum water content: 7 lb/MMSCF
Condensable hydrocarbon content: 2 US gallon/MMSCF (maximum**)**

3 SIZING OF THE EQUIPMENT

3.1 INLET 3-PHASE SEPARATOR

Liquid capacity of separator
= (1.65 bbl/MMSCF×40 MMSCF/day)=66 bbl/day.

The OD, height and settling volume of the separator can be determined *[from Appendix B, Table 1]* and given bellow:

Size of OD (outer diameter) =16''
Shell height =7.5'
Settling volume = 0.72 bbl

3.2 GLYCOL CONTACTOR

A Valve Tray type Glycol Contactor Tower will be designed under the operating condition.
A Glycol Contactor diameter is selected based on the operating pressure required with the approximate required gas capacity.
For 40 MMSCFD and 1000 psig working pressure, contactor size can be found. *[from Appendix B, table 4]*
Size of OD (outer diameter) = 48''
Gas capacity of contactor at standard condition, q_S = 49.6 MMSCFD
(Standard condition means at 0.7 sp. gravity and 100°F)
Now, two correction factors are needed to be introduced.
C_t = Temperature correction factor
C_g = Sp. Gravity correction factor
Now, at 120^{0}F and gas with specific gravity of 0.6.
We find, C_t = 0.98 and C_g = 1.08 *[from the Appendix B, Table 2 & 3]*
So, Gas capacity of contactor at operating condition, $q_0 = q_s \times C_t \times C_g$
So, q_o = 52.4 ≅ 52 MMSCFD.
So, a 48 inch outer diameter contactor cannot handle up to 52 MMSCFD gas capacity.
So, the following data can be found. *[from Appendix B, Table 4]*
Size of OD = 54 inch
Gas inlet and outlet size = 6 inch
Glycol inlet and outlet size = 2 inch
Glycol Cooler Size = 6 inch×8 inch
Shipping weight = 20200 lb

Calculation of number of trays in the contactor:

Gas flow rate = 40 MMSCFD
Gas sp. Gravity = 0.6
Operating Pressure = 1000 psig
Gas inlet temperature = 120°F
Gas outlet water content = 7 lb H_2O/MMSCF

The following data are listed below: *[From Appendix A, Figure 1]*

Table-1: Dew Point Depression and Water Content calculation

	Dew Point Temperature (^{0}F)	Water Content (lb H_2O/MMSCF)
Inlet	120	100
Outlet	32	7
Dew Point Depression	88	93

Now, C_{rich} = (0.995×9.2)/(9.2+1/3.5) = 96.50%

Where, C_{lean} = 99.5%, L_w = 3.5gal/lb H_2O removed

P_{lean} = (1.103×8.34) lb_m TEG/gal *[from Appendix B, Table 11]*

Now, Number of actual trays = 6.8 ≅ 7 *[from Appendix A, Figure 14]*

Here, the operating line points are given below:

Top of Column: 7 lb H_2O/MMSCF and 99.5% TEG

Bottom of Column: 100 lb H_2O/MMSCF and 96.55% TEG.

Table-2: Equilibrium line calculation

% TEG	Equilibrium dew point temp. at 120^{0}F *[from Appendix A, Figure 2]*	Water content of natural gas at dew point temp. and 1000 psig from *[Appendix A, Figure 1]* **(lb H_2O/MMSCF)**
99.6	13	3.05
99	37	8.5
98	53	13.5
97	60	17.5
96	70	23
95	74	27.4

Now, to determine the triangulation, the following graph is plotted.

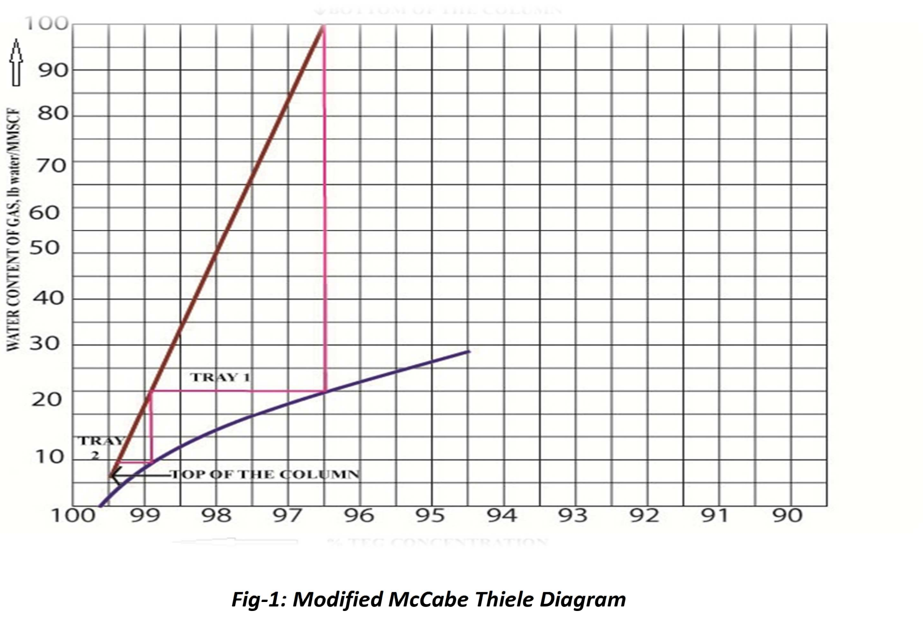

Fig-1: Modified McCabe Thiele Diagram

So, from Modified McCabe-Thiele diagram, theoretical number of trays required is 2.8.

Now, Number of actual trays = 2.8/0.33 = 8.48 ≅ 9 (assume, valve trays)

Standard field dehydration contactors normally have 24 inch spacing.

So, Number of actual trays = 9

3.3 LEAN-RICH HEAT EXCHANGER

Assume a 1-2 pass shell and tube heat exchanger where hot fluid flows in the shell and cold fluid flows in the tube.

Here, hot fluid is lean glycol and cold fluid is rich glycol.

Now, heat balance =207383.68 Btu/hr

Amount of hot fluid = (33.6×150.17+1.41×18) lb_m/hr = 5071.09 lb_m/hr

Amount of cold fluid = (33.6×150.17+10.022×18) lb_m /hr =5226.108 lb_m /hr

So, LMTD = 101.93^0F

Correction of LMTD:

Temperature effectiveness = 0.43

Temperature difference ratio = 0.94

LMTD correction factor, F_T = 0.94 *[from Appendix A, Figure 7]*

So, Corrected LMTD = previous LMTD× F_T = 96°F

Now,

Sp. gravity of hot fluid (lean glycol) at 350°F is 1.005 ≅ 1.0 *[from Appendix A, Figure 6]*

So, API^0 = 141.5/1-131.5=10^0 API

And,

sp. Gravity of cold fluid (rich glycol) at 175.95 ^{0}F is 1.06 *[from Appendix A, Figure 6]*

So, API^0 = 141.5/1.06-131.5=2^0 API

If, the viscosities of both hot and cold fluid don't vary too much. Then the caloric temperature can be determined from the mean temperature. Though the both fluid is TEG, so the caloric temperature can be determined by following:

Caloric temperature of hot fluid. T_c = 315°F

Caloric temperature of cold fluid, t_c= 213°F

Assumption of U_D

Now, the viscosities at both calorific values are given here. *[from Appendix A, Figure 5]*

For 99.5% TEG, at 315°F; viscosity of hot fluid, μ_h= 0.8 cp

For 96.5% TEG, at 213°F; viscosity of cold fluid, μ_c= 3.10 cp

Now, the cold fluid passes in the tube. So, U_D will be assumed for the viscosity of cold fluid.

For, μ_c =3.10 cp, the range of the U_D = 10 to 40 Btu/hrft^2°F. *[from Appendix B, Table 5]*

Selection of Tube length and Tube layout

A cost variation curve to select the length of tube is given here.

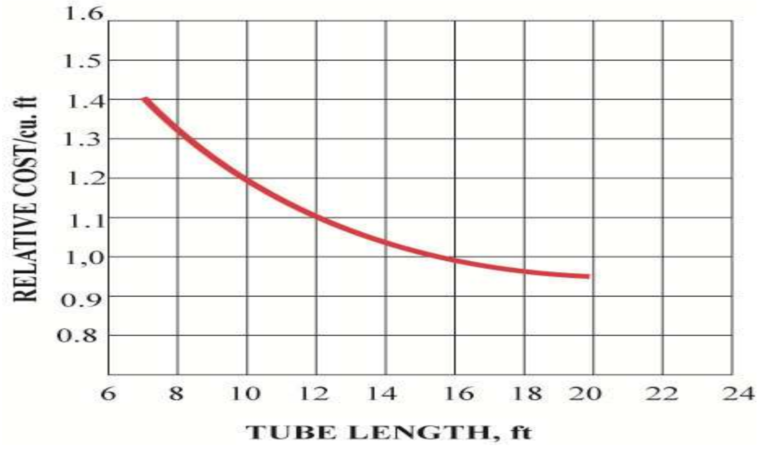

Fig-2: Cost of Tubular surface vs tube length [After D. Q. Kern, 1983]

From the figure, cost variation between the use of 12 ft, 16 ft and 20ft is not very great.

So, we assume the tube length is 12 ft

The advantage of a square pitch over triangular pitch is the tubes are accessible for external cleaning and cause a lower pressure drop when fluid flows in the direction.

A cost variation curve to select the tube OD is given here.

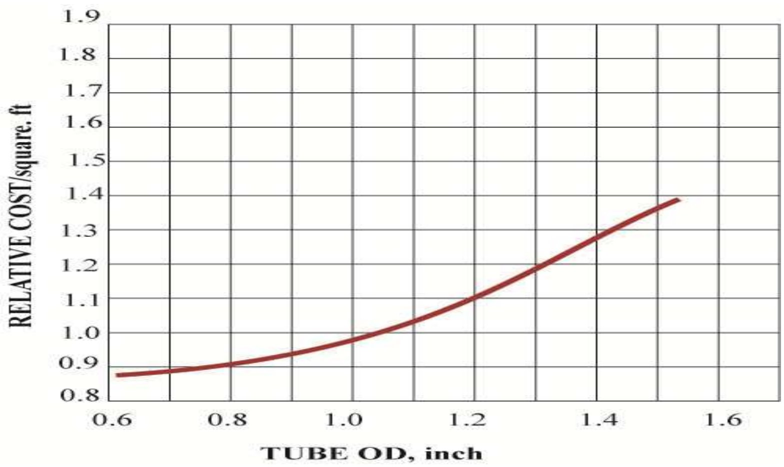

Fig-3: Cost of Tubular surface vs tube OD [After D. Q. Kern, 1983]

From the figure it is clear that the cost for ¾ inch OD is relatively low and feasible than the others. So, a OD of ¾ inch can be selected.

So. Square tube pitch selection is appreciable. Most commonly used pitches for square layout is ¾ inch. OD on 1 inch square pitch.

Generally, 10, 12,14 or 16 BWG are available selection. We assume a 14 BWG for heat exchanger tube

Determining clean overall heat transfer co-efficient and dirt factor

Now, for a tube length of 12ft, ¾ inch. OD on 1 inch square pitch and 14 BWG we get following values. *[from Appendix B, Table 7]*

Tube ID = 0.584 in

Flow area per tube = 0.268 in^2

Surface per lin ft = 0.1963 in^2

Tube Length = 12 ft

By Trial and error method for different $U_{D,}$ clean overall heat transfer co-efficient is determined.

Methodology

- $A=Q/(U_D\ \Delta t)$
- Number of tubes=A/(L×a''),
- Select a nearest number of tubes*. [from Appendix B, Table 7]*
- Determine corrected U_D

Cold Fluid (Rich Glycol): Tube Side

- Obtain $\mu_{cold\ actual}$ at t_c *[from Appendix A, Figure 5 & 8]*
- Obtain sp. Gravity (API^0) at t_c *[from Appendix A, Figure 6]*
- $a_t\ (ft^2) = (N_t \times a'_t)/(144\times n)$
- $G_t\ (lb/(hr)(ft^2))=W/a_t$
- $R_{et} = D\ G_t/\ \mu_{cold\ actual}$
- Obtain J_H for R_{et} and (L/D) *[from Appendix A, Figure 9]*
- Obtain k((cμ)/k)^(1/3) for $\mu_{cold\ actual}$ (cp) and sp. gravity (API^0) *[from Appendix A, Figure 10]*
- $H_{to} = J_H\times(1/D)\times k((c\mu)/k)^{(1/3)}\times\phi_t$
- $H_t = h_{so}\times ID/OD$

Hot Fluid (Lean Glycol): Shell Side

- Obtain $\mu_{hot\ actual}$ at T_c *[from Appendix A, Figure 5 & 8]*
- Obtain Sp. Gravity (API^0) at T_c *[from Appendix A, Figure 6]*
- B = Shell ID/5
- $a_s\ (ft^2)=(ID\times BC')/(144\times P_T)$
- $G_t\ (lb/(hr)(ft^2))=W/a_s$
- Obtain D_e *[from Appendix A, Figure 11]*
- $R_{es} = D_e\ G_t/\mu_{hot\ actual}$
- Obtain J_H *[from Appendix A, Figure 11]*
- Obtain k((cμ)/k)^(1/3)for $\mu_{hot\ actual}$ (cp) and sp. gravity (API^0) *[from Appendix A, Figure 10]*
- $h_s=J_H\times(1/\ D_e)\times k((c\mu)/k)^{(1/)3}\times\phi_s$

- Clean Overall Heat Transfer Co-efficient,

  $U_C=(h_s\times h_t)/(h_s+h_t)$
- Trial and error method will continue until $U_C>U_D$

Overall Clean Heat Transfer Co-efficient is slightly lower than the range which is So, U_C = 6.53 Btu/hr.^{0}F.ft^2 (at U_D = 6Btu/hr.^{0}F.ft^2 and corrected U_D = 4.9 Btu/hr.^{0}F.ft^2)

Pressure Drop Calculation

Pressure drop can be calculated using the following method.

Cold Fluid (Rich Glycol): Tube Side

For R_{et} = 173, *[from Appendix A, Figure 12]*

Friction factor, f = 0.0025

Sp. Gravity, s = 1.06

So, Pressure Drop

$\Delta P_t = f\ G_t^2\ Ln/5.22\times10^{10}\ D_t\ s\phi_t$

=0.0145 psi

Hot Fluid (Lean Glycol): Shell Side

For R_{es}=1600, *[from Appendix B, Figure 13]*

Friction factor, f = 0.003

Sp. Gravity, s = 1

Now,

No. of crosses, N+1=12L/B= 36

D_s = 19.25/12=1.604

So, Pressure Drop

$\Delta P_s = f\ G_s^2\ D_s\ (N+1)/5.22\times10^{10}\ D_{es}$

=0.0637 psi

Dirt Factor (R_d) Calculation:

$R_d = (U_C - U_D)/(U_C\ U_D)$ = 0.0509 hr.^{0}F.ft^2/Btu

So, dirt factor is 0.0509 hr.^{0}F.ft^2/Btu.

Table-3: Summary of the heat exchanger design

SHELL SIDE		TUBE SIDE
45.19	$H_{Outside}$	7.63
U_C		6.53
U_D		4.9
R_D Calculated		0.0509
0.0637	Δp Calculated	0.0145

Final selection of exchanger

Tube side:

Number of tubes =220

BWG =14

OD = ¾ inch

Length = 12 feet

Pitch: 1 inch square

Passes: 2

Shell side:

ID = 19.25 inch

Pass: 1

3.4 GLYCOL CIRCULATION PUMP

The required size of glycol circulation pump can be readily determined using the glycol circulation rate and maximum operating pressure of the contactor.

Water in flowing gas = 100 lb /MMSCF (at 120^0 F and 1000psia)

Glycol circulation required = 3.5 gallon TEG/lbH_2O

Circulation Rate = (3.5galTEG/1lbWater)×(100lbWater/1MMScf)

× (1day/24hr) × (1hr/60min)×(40MMScf/1day)= 9.7 GPM

Table-4: Comparison between Union Pump and KIMRAY Pump [Titas Gas Field data]

	Union Pump	**Proposed KIMRAY Pump**
Capacity (max.)	17.6 GPM	7.5 GPM
Speed (max)	50 SPM	28 SPM
Per Stroke	0.352 G	0.2678 G

So, to reduce the cost it is recommended to use the proposed KIMRAY pump even the capacity is lower than we needed. From Titas Gas field data, it is recommended to select KIMRAY Pump & model number is 45015 PV. *[Titas Gas Field (Location #A) Data]*

3.5 CONDENSATE FLASH SEPERATOR (3-PHASE)

A flash separator should be installed downstream from the glycol pump to remove any entrained hydrocarbon from the rich glycol. The separator is sized based on a liquid retention time in the vessel of at least 5 minutes.

The size can be calculated by following equation: *[After Sanjay Kumar, 1987, vol 4]*

V=Lt/60

Where,

V= Settling volume required in the separator, gal

t = Liquid retention time, min

V = 583.33×5/60= 48.61 gal ×1bbl/(42 gal)=1.15 bbl

Now, the following values can be found. *[from Appendix B, Table 1]*

Size of OD = 20 inch

Shell height = 7.5 ft

3.6 GLYCOL REGENERATOR (RE-BOILER)

The glycol circulation rate in Gallons per Hour L, in the plant is given by following equation:

[After Sanjay Kumar, 1987, Vol 4]

$L=L_w W_i (q/24)$

L= 3.5×100×(40/24) = 583.33 gal/hr

An approximate calculation of the heat required in the re-boiler also called re-boiler heat load, Q in Btu/hr, can be made using the following empirical relationship:

Q = 2000L=2000×583.33 = 1166660 BTU/hr

So, for the heat load of the re-boiler is 1166660BTU/hr, the design for the re-boiler can be found.

[from Appendix C, Table 8]

3.7 STRIPPING STILL

The diameter (or cross sectional area) of the packed stripping still for use with the glycol re-concentrator can be estimated,*[from Appendix A, Figure 3]* as a function of the glycol re-concentrator's glycol to water circulation rate gal TEG/lb water and the glycol circulation rate (gal/hr). The size of a stripping still is governed by vapor and liquid loading conditions at its bottom. The vapor load comprises vapor load and stripping gas flowing upwards through the still, whereas the liquid load consists of rich glycol and the reflux flowing downwards. Generally stripping gas requirements, in the range of 2-10 ft^3/gal TEG circulated. *[Appendix A, Figure 3]*

For a packed type design, a minimum of 4 feet packing height is provided, consisting generally of 1.5 inch ceramic, saddle type packing. This height should be increased with increasing glycol re-concentrator size, to a maximum of about 8 ft for a 1 MM BTU/hr. *[After Sanjay Kumar, 1987, vol 4]*

For a glycol circulation rate of 583.33 gal TEG/hr, re-boiler heat load of 1.167×10^6 Btu/hr and with intalox saddles (1.5 inch) type packing. Now, the size of the stripping still can be given following: *[from Appendix B, Table 8]*

Stripping column diameter = 16 inch

Stripping column height = 8 ft

The amount of stripping gas required to re-concentrate the glycol to a high percentage is 6 ft^3/gal of glycol circulated i.e. the amount is 3499.98 ft^3/hr

4 Discussions & Recommendations

This paper work is based on the operating condition of Titas Gas Field (Location #A). Here operating conditions are the gas flow rate, wellhead temperature, pressure, gas composition, gas properties, water-content in raw gas etc. So, with the change of these conditions the total results can be changed also. This project work is mainly a model to design any glycol dehydration plant. The results we got in the overall paper are established on theoretical proof and some practical knowledge. So, the procedure to design any glycol dehydration plant will be similar as we have designed here but the results can be varied.

In the sizing of Inlet 3-phase separator the liquid capacity is 66 bbl/day. But, the OD and height of the separator is designed for the liquid capacity of maximum 100bbl/day to satisfy any overflow of gas. Though the operating condition is not the standard condition, so in the sizing of glycol contactor two correction factors of temperature and sp. gravity are introduced to determine gas capacity. And the contactor is designed for the maximum gas capacity of 52 MMSCFD. In the lean-rich heat exchanger, selection of tube length, tube layout, BWG can be varied which may affect the value of clean and dirt overall co efficient and pressure drop as well as the dirt factor. The glycol circulation rate generally exceeds the maximum circulation rate for the proposed KIMRAY pump of model number 45015PV. Theoretically this model is not recommended to use, but practically this pump is used for economic feasibility. For a general glycol regenerator, the maximum heat load is 1000000 Btu/hr. But, the heat load in the re-boiler we got in this process exceeds the maximum value. Still the re-boiler and stripping still is designed according to the maximum heat load of 1000000 Btu/hr.

This paper work is a model to design any glycol dehydration process plant. As, the project work is based on the operating condition of Titas Gas Field (Location #A) where the raw natural gas is sweet gas and water is the major impurity. So, this work is recommended to follow for those gas fields where the properties of raw natural gas are similar to the Titas Gas Field. The work is done for a single well of 40 MMSCFD and can be elaborated for multiple wells in a single process plant of higher capacity.

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6 Nomenclature

^0F = Degree Fahrenheit
μ_c = Viscosity of Cold Fluid
$\mu_{hot\ actual}$ = Viscosity of hot fluid
$\mu_{cold\ actual}$ = Viscosity of cold fluid,
ϕ_s = Viscosity ratio for shell side
ϕ_t= Viscosity ratio for tube side
h_s = Heat transfer co-efficient for shell
h_t= Heat transfer co-efficient for tube
ID = Inner Diameter
J_H= Factor for heat transfer
k = Permeability
L = Glycol circulation rate, gal/hr

ΔT = Corrected LMTD, ^0F
ΔT = Temperature Difference, ^0F
a_s= Flow area
a_t= Actual flow area
A = Total Tube Area, ft^2
a'_t= Flow area
a" = Surface Perlinft, in^2
B = Maximum baffle spacing,
bbl = Barrel
C_g = Sp. Gravity correction factor
C_p = Specific Heat, BTU/lb^{0}F
C_t = Temperature Correction Factor
C'= Clearance
$D_e = d_e$ /12, d_e = Equivalent diameter
d_t = Tube ID
D = Nominal OD, inch
D = d_t/12,
Et al = Associates
Etc. = Etcetera
F = Design Factor
G_s= Mass velocity
G_t= Mass velocity
G_D = Dry gas
L_w = Glycol to water circulation rate, gal/lb H_2O
L= Tube length, ft
LMTD = Logarithmic Mean Temperature Difference
m = Amount of Glycol, lb_m/hr
MMSCF = Million Standard Cubic Feet
n = Water content in glycol, lb_m/hr
n= Number of tube pass
N_t = Number of tubes
OD = Outer Diameter
P_T = Number of shell pass
P= Design Pressure, psi
Q = Amount of Heat, Btu/hr
q = Gas flow rate, MMSCFD
q_o = Gas capacity of contactor, MMSCD
R_{es} = Reynolds number
R_{et}= Reynolds number
t = Liquid retention time, min
T = Temperature, ^0F
t = Wall Thickness, inch
T_c = Calorific Temperature of Hot fluid, ^0F
t_c= Calorific Temperature of Cold fluid, ^0F
U_D = Overall Dirt Co-efficient, BTU/hr^{0}Fft^2

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