American Institute
Of
Chemical Engineers

NATIONAL STUDENT DESIGN COMPETITION

1991

AIChE

345 East 47 Street • New York, New York 10017

NOTES

MATERIAL BALANCE, GAS CONCENTRATION UNIT, OPERATIONS 1990

STREAM LB MOL/HR COMPONENT, MOL WT		(25) SPLITTER REBOILER CIRCULATN	(26) BLEND BUTANE TO STORAGE
HYDROGEN	2.016		
CO2	44.010		
H2S	34.080		
METHANE	16.043		
ETHANE	30.070		
PROPANE	44.097		
I-BUTANE	58.124	110.20	8.60
BUTANE	58.124	2780.30	215.80
I-PENTANE	72.151	362.30	28.10
PENTANE	72.151	226.10	17.50
HEXANE	86.178	0.20	TR
NONANE	128.260		
DECANE	142.287		
UNDECANE	156.314		
TOTAL		3479.10	270.00
TEMPERATURE F		157.00	100.00
PRESSURE, PSIA		105.00	98.00
ENTHALPY, MILLION BTU/HR		-1346.00	-17.00
VAPOR, MOL %		0.00	0.00
MASS FLOW, LB/HR		210480.00	16334.00
STD LIQ VOL, BBL/HR		1016.00	79.00
DENSITY, LB/CU FT		32.80	35.40
MOLECULAR WEIGHT		60.50	60.50
HEAT CAPACITY, BTU/LB DEG F.		0.68	0.61

AIChE NATIONAL STUDENT DESIGN COMPETITION
1991
Revamp Refinery NGL Processing Unit

DEADLINE FOR MAILING

Solution must be postmarked not later than midnight, June 3, 1991

RULES OF THE CONTEST

Solutions will be graded on (a) substantial correctness of results and soundness of conclusions, (b) ingenuity and logic employed, (c) accuracy of computations, and (d) form of presentation. Accuracy of computations is intended to mean primarily freedom from mistakes; extreme precision is not necessary.

It is to be assumed that the statement of the problem contains all the pertinent data except for those readily available in handbooks and similar reference works. The use of textbooks, handbooks, journal articles, and lecture notes is permitted. In cases where there is disagreement in the data reported in the literature, the values given in the problem have been chosen as being most nearly applicable.

Students may use any available commercial or library computer programs in preparing their solutions. Students are warned, however, that physical property data built into such programs may differ from data given in the problem statement. In such cases, as with data from other literature sources, values given in the problem statement are most applicable. Students using commercial or library computer programs or other solution aids should so state in their reports and include proper references and documentation. Students are further advised that the problem can be solved without the use of sophisticated computer programs. Judging is based on the overall suitability of the solution, not on skills in manipulating computer programs.

The Student Contest Problem is designed to be solved by individual chemical engineering students working entirely alone, and it is judged on that basis. There are, however, other academically sound approaches to using the problem. The following confidentiality rules therefore apply:

1. ***For students whose solutions may be considered for the contest:***
   The problem may not be discussed with anyone (students, faculty, or others, in or out of class) before or during the period allowed for solution. Discussion with faculty and students at that school is permitted only after complete final reports have been submitted to the chapter counselor.

2. ***For students whose solutions are not intended for the contest:***
   Discussion with faculty and with other students at that school who are not participating in the contest is permitted.

3. ***For all students:***
   The problem may not be discussed with students or faculty from other schools, or with individuals in the same school who are still working on the problem for the contest, until after June 3, 1991. This is particularly important in cases where neighboring institutions may be using different schedules.

Submission of a solution for the competition implies strict adherence to these conditions.

A period of not more than thirty days is allowed for completion of the solution. This period may be selected at the discretion of the individual counselor, but in order to be eligible for an award a solution must be postmarked not later than midnight, June 3, 1991. ONLY SOLUTIONS SUBMITTED BY NATIONAL STUDENT MEMBERS OF AIChE WILL BE CONSIDERED FOR AWARDS.

The finished report should be submitted to the chapter counselor within the thirty-day period. There should not be any variation in form or content between the solution submitted to the chapter counselor and that sent to the AIChE office. The body of the report must be suitable for reproduction, that is, typewritten or computer-generated. Tables may be written in ink. Supporting calculations and other appendix material may be in pencil. Each counselor should select the best solution or solutions, not to exceed two, from his or her chapter and send these by registered mail to the institute.

Two copies of the solution(s) must be accompanied by a letter of transmittal giving only the contestant's name, school address, home address, home telephone number, and student chapter, lightly attached to the report. This letter will be retained for identification by the executive director of the Institute. The solution itself must bear no reference to the student's name or institution by which it might be identified. In this connection, graph paper bearing the name of the institution should be avoided. Original manuscript(s) must remain in the possession of the student chapter counselor, or faculty member, sponsoring the student(s).

As soon as the winners have been notified, original manuscripts for first, second, third and honorable mention categories must be forwarded to the office of the Executive Director as soon as possible.

Richard E. Emmert
Executive Director
American Institute of Chemical Engineers
345 East Forty-seventh Street
New York, New York 10017

Student chapter counselors and design course instructors are asked to contact Dr. Jerald N. Linsley, Chair of the 1991 National Student Competition Subcommittee if they have any questions about the design problem.

Dr. Jerald N. Linsley
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Melbourne, FL 32901-6988

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*Dial 407/768-8000 and ask for extension 7560

PLEASE READ THE RULES CAREFULLY
BEFORE SUBMITTING A SOLUTION TO AIChE

American Institute of Chemical Engineers, Inc.
345 East Forty-seventh Street
New York, New York 10017

MATERIAL BALANCE, GAS CONCENTRATION UNIT, OPERATIONS 1990

STREAM LB MOL/HR COMPONENT, MOL WT		(19) REFLUX TO DeC3	(20) PROPANE TO DEH & STG	(21) BOTTOM FROM DeC3	(22) FEED TO C4 SPLITTER	(23) REFLUX TO C4 SPLITTER	(24) IC4 TO C4 SPLITTER
HYDROGEN	2.016						
CO2	44.010						
H2S	34.080						
METHANE	16.043						
ETHANE	30.070	6.80	0.60				
PROPANE	44.097	587.40	54.80	2.20	2.20	20.30	2.20
I-BUTANE	58.124	6.00	0.60	171.10	171.10	1481.10	162.50
BUTANE	58.124	0.10	TR	226.60	226.60	98.60	10.80
I-PENTANE	72.151			28.10	28.10		
PENTANE	72.151			17.50	17.50		
HEXANE	86.178			TR	TR		
NONANE	128.260						
DECANE	142.287						
UNDECANE	156.314						
TOTAL		600.30	56.00	445.50	445.50	160.00	175.50
TEMPERATURE F		101.00	101.00	229.00	143.00	101.00	101.00
PRESSURE, PSIA		337.00	337.00	280.00	102.00	142.00	142.00
ENTHALPY, MILLION BTU/HR		-30.60	-2.80	-26.60	-27.90	104.40	-11.50
VAPOR, MOL %		0.00	0.00	0.00	11.00	0.00	0.00
MASS FLOW, LB/HR		26462.00	2466.00	26510.00	26510.00	92713.00	10176.00
STD LIQ VOL, BBL/HR		149.00	14.00	130.00	130.00	470.00	51.00
DENSITY, LB/CU FT		29.50	29.50	27.70	-	33.40	33.40
MOLECULAR WEIGHT		44.10	44.10	59.50	59.50	57.90	57.90
HEAT CAPACITY, BTU/LB DEG F.		0.73	0.73	0.87	0.69	0.62	0.62

MATERIAL BALANCE, GAS CONCENTRATION UNIT, OPERATIONS 1990

STREAM	(13)	(14)	(15)	(16)	(17)	(18)
LB MOL/HR	REFLUX	NGL	BOTTOMS	DEC5	DEC5	FEED
	TO	TO	FROM	REBOILER	OIL TO	TO
COMPONENT, MOL WT	DEPENTANZR	TREAT	DEPENTANZR	CIRCULATN	HDS	DEPROPANZR
HYDROGEN 2.016						
CO2 44.010						
H2S 34.080	TR	0.10				
METHANE 16.043						
ETHANE 30.070	0.40	0.60				0.60
PROPANE 44.097	33.30	57.00				57.00
I-BUTANE 58.124	100.50	171.70	0.10	0.60	0.10	171.70
BUTANE 58.124	132.70	226.60	0.70	3.50	0.40	226.60
I-PENTANE 72.151	16.50	28.10	11.90	61.30	6.60	28.10
PENTANE 72.151	10.30	17.50	16.10	83.10	8.90	17.50
HEXANE 86.178	TR	TR	27.50	142.30	15.30	TR
NONANE 128.260			57.00	294.70	31.80	
DECANE 142.287			104.70	541.40	58.40	
UNDECANE 156.314			16.10	83.20	9.00	
TOTAL	293.70	501.60	234.10	1210.10	130.50	501.50
TEMPERATURE F	100.00	100.00	410.00	410.00	106.00	171.00
PRESSURE, PSIA	105.00	105.00	114.00	114.00	33.00	294.00
ENTHALPY, MILLION BTU/HR	-18.70	-31.90	-21.10	-108.80	-14.80	-30.50
VAPOR, MOL %	0.00	0.00	0.00	0.00	0.00	0.00
MASS FLOW, LB/HR	16967.00	28979.00	29161.00	150762.00	16257.00	28976.00
STD LIQ VOL, BBL/HR	85.00	144.00	116.00	600.00	65.00	144.00
DENSITY, LB/CU FT	34.20	34.20	33.60	33.60	43.70	30.90
MOLECULAR WEIGHT	57.80	57.80	124.60	124.60	124.60	57.80
HEAT CAPACITY, BTU/LB DEG F.	0.62	0.62	0.68	0.71	0.54	0.71

Revamp Refinery NGL Processing Unit

Introduction

A small refinery serves a local Western U.S. market for motor fuel, diesel fuel, propane and asphalt. Recently completed process changes have increased the demand for blending butanes (iso-butane and normal butane). This increased demand can be satisfied by either processing additional natural gas liquids (NGL) or by importing purified butanes by rail car. A recent discovery in the area promises increased availability of natural gas liquids. For negotiating purposes, the refinery management must have plant modification requirements, plus capital and operating cost figures, for processing the additional NGL.

Economic Information

Butane pricing is as follows:

C3	(fuel)	20 cents per gallon
isobutane	(blend stock)	55 cents per gallon
n-butane	(blend stock)	45 cents per gallon
C5+	(blend stock)	55 cents per gallon
n-butane	(sales)	30 cents per gallon
isobutane	(import by rail)	45 cents per gallon
n-butane	(import by rail)	35 cents per gallon

The refinery requires a 20 percent return on capital investment.

Problem Statement

1) Determine the technical and economic feasibility of importing the NGL for the two case listed below.

   Case 1 - Replacement of the normal butane rail car imports.
   Case 2 - Replacement of both the isobutane and normal butane rail car imports.

2) Determine the maximum price which can be paid for the field butanes and still meet the refinery investment criteria for both cases.

KICK-OFF MEETING

SUBJECT: NGL Kickoff Meeting

LOCATION: Refinery Conference Room

PRESENT: Refinery Manager and Department Supervisors

Meeting was opened by Bill, Refinery Manager, who stated that recent drilling activity in the area has been successful and will result in additional quantities of NGL being available. In order to negotiate with the supplier to purchase these additional NGL's we need to know how much is needed and can be handled through the refinery facility.

Logistics Supervisor, Lonnie, showed the following information:

Current NGL Purchases	600,000 Bbls/Year
Expected Additional N-Butane Requirements	10,000 Bbls/Year
Expected Additional I-Butane Requirements	35,000 Bbls/Year

Lonnie also stated that current handling of blending butane requires seasonal storage and special handling because the blending butane does not meet accepted industry standards. While it is expected that motor fuel blending would require an additional 10,000 barrels of blending butane, quantities in addition to that would have to meet industry standards for N-Butane. Lonnie stated that when renegotiating the NGL contract, a quality specification should be imposed as he is not comfortable bringing any material into the refinery without specifications. His suggestion was to limit vapor pressure to 70 psia at 100°F, prohibit free water and require current supply pressure to be maintained. The composition of the imported NGL is anticipated to be identical to the current NGL charge, stream 4 of the material balance.

MATERIAL BALANCE, GAS CONCENTRATION UNIT, OPERATIONS 1990

STREAM LB MOL/HR COMPONENT, MOL WT	(7) LEAN OIL TO DEETHANIZR	(8) VAPOR FEED TO DEETHANIZR	(9) LIQUID FEED TO DEETHANIZR	(10) DEETHANIZR O H VAP TO FUEL	(11) DEETHANIZR BTM FM REBOILER	(12) FEED TO DEPENTANZR
HYDROGEN 2.016		34.70	2.40	37.10		
CO2 44.010		0.20	0.20	0.40		
H2S 34.080		0.30	0.60	0.80	0.10	0.10
METHANE 16.043		37.80	13.50	51.30		
ETHANE 30.070		13.60	21.60	34.50	0.60	0.60
PROPANE 44.097		10.10	48.50	1.70	57.00	57.00
I-BUTANE 58.124	TR	14.70	157.10	TR	171.80	171.80
BUTANE 58.124	0.30	14.70	212.30	0.10	227.30	227.30
I-PENTANE 72.151	5.20	1.00	34.10	0.40	40.00	40.00
PENTANE 72.151	7.10	0.60	26.40	0.50	33.60	33.60
HEXANE 86.178	12.20	0.10	15.50	0.30	27.50	27.50
NONANE 128.260	57.00			0.10	57.00	57.00
DECANE 142.287	104.70			TR	104.70	104.70
UNDECANE 156.314	16.10			TR	16.10	16.10
TOTAL	202.80	127.90	532.10	127.10	735.70	735.70
TEMPERATURE F	105.00	99.00	99.00	112.00	227.00	232.00
PRESSURE, PSIA	264.00	217.00	217.00	202.00	205.00	130.00
ENTHALPY, MILLION BTU/HR	-24.20	-4.00	-33.30	-3.10	-54.10	-51.40
VAPOR, MOL %	0.00	100.00	0.00	100.00	0.00	48.00
MASS FLOW, LB/HR	26702.00	3392.00	30208.00	2163.00	58139.00	58139.00
STD LIQ VOL, BBL/HR	105.00	24.00	152.00	20.00	260.00	260.00
DENSITY, LB/CU FT	44.20	1.03	34.10	0.57	33.90	-
MOLECULAR WEIGHT	131.70	26.50	121.40	17.00	98.30	98.30
HEAT CAPACITY, BTU/LB DEG F.	0.53	0.53	0.62	0.60	0.67	-

VESSELS

V-1 DE-ETHANIZER ABSORBER

Bottoms Sect. - 4'6" dia. (trays 1 thru 23)
Top Section - 3' dia. (trays 24 to 40)
40 Actual Trays - 16 theoretical including reboiler
Tray Spacing - 24"
Overall Height - 89'6" TT
Design Conditions: 229 psig, 235°F

V-2 DEPENTANIZER

4' dia. X 74' overall height TT
30 Actual Trays - 22 theoretical incl. condenser & reboiler.
Tray Spacing - 24"
Design Conditions: 120 psig, 650°F

V-3 DEPENTANIZER REFLUX ACCUMULATOR

7' dia. X 21' TT horizontal
Design: 120 psig, 650°F

V-4 LEAN OIL SURGE TANK

7' dia. X 16' TT horizontal
Design: 60 psig, 650°F

V-5 COMPRESSOR - FIRST STAGE SUCTION SCRUBBER

3'6" dia X 7' TT vertical
Design: 103 psig, 650°F

V-6 COMPRESSOR - SECOND STAGE SUCTION SCRUBBER

3'6" dia X 9' TT vertical
Design: 103 psig, 650°F

V-7 CAUSTIC TREATER

7'6" dia X 24' TT vertical
Design: 360 psig, 300°F

3. VESSELS

EXISTING AND NEW

Use six (6) minutes minimum hold-up, based on liquid pump-out rate. The diameter is to be divisible by 0.5 feet. (rev. 1)

Vapor-liquid capacity equation applies to single pass valve trays with 15% downcomers and 24 inch tray spacing.

rev 1

$$V_{load}\ (design) = D^2\ (0.257 - .00141\ \frac{GPM}{D}) \quad \text{for } \frac{GPM}{D}\ 30\text{-}100^*$$

$$V_{load} = \text{Vapor Rate, } Ft^3/sec * (Dv/(Dl-Dv))^{1/2}$$

D = Diameter, ft.
Dv = Vapor density lb/ft^3
Dl = liquid density lb/ft^3

* If liquid flow is less than 30 GPM/foot use 30. In columns where the feed rate is not directly controlled by the operator, limit load to 90% of normal design. The minimum diameter for trays is 3'-0". (rev 1)

NEW Equations useful to estimate thickness of vertical carbon steel vessels:

$$t_h = \frac{PD}{2479-.1P}$$

(round up to next 1/16 inch)
P = Design Pressure, psig
D = Diameter, ft
t_h = Thickness req'd for hoop stress in inches

$$t_a = \frac{90 \times 10^{-6}\ (H+S)^2}{D}$$

(round up to next 1/16 inch)
H = Vessel Height, ft
S = Skirt Height, ft
t_a = Thickness req'd for wind velocity in inches

$t = t_h + 1/8$ for t_h greater than t_a

$t = 0.5(t_h + t_a) + 1/8$ for t_h less or equal to t_a

Insulation for pipes use 1" minimum to protect personnel
Insulation for Vessels use 1" for 100°F,
4" for 300°F, interpolate and round up to next 1/2".

4. **PIPING**

<u>EXISTING</u> Existing pipes are sized conservatively and will handle 20% more flow without excess pressure drop.

<u>NEW</u> Use standard pipe sizes for pipelanes 2, 4, 6, 8 and 10. Size for 4-7 ft/sec where possible.

5. **MISCELLANEOUS**

<u>MAINTENANCE</u> = 3% of investment/year 40% labor
60% materials.

<u>TECHNICAL & SUPERVISION</u> = 25% of Operating & maint. labor

<u>PUMPS</u>

P-1, P-2, P-3 DE-ETHANIZER INTERCOOLER PUMPS

Each: 190 GPM, 100 ft head, Sp. Gr. 0.61 @ pumping temperature of 138^{o}F.

P-4 DEPENTANIZER REBOILER PUMP

764 GPM, 300 ft head, Sp. Gr. 0.51 @ 463^{o}F.

P-5 DEPENTANIZER REFLUX PUMP

70 GPM, 200 ft. head, Sp. Gr. 0.52 @ 100^{o}F.

P-6 LEAN OIL PUMP

75 GPM, 803 ft. head, Sp. Gr. 0.78 @ 100^{o}F.

P-7 COMPRESSOR SCRUBBER LIQUIDS PUMP

20 GPM, 920 ft. head, Sp. Gr. 0.56 W 100^{o}F.

P-8 DEPROPANIZER REFLUX PUMP

195 GPM, 325 ft. head, Sp. Gr. 0.26 @ 100^{o}F.

P-9 BUTANE SPLITTER REFLUX PUMP

475 GPM, 635 ft. head, Sp. Gr. 0.54 @ 100^{o}F.

P-10 BUTANE REBOILER CIRCULATING PUMP

800 GPM, 150 ft. head, Sp. Gr. 0.526 @ 175^{o}F.

P-11 DEPROPANIZER FEED PUMP

130 GPM, 1,240 ft. head, Sp. Gr. 0.52 @ 100^{o}F.

E-14 DEPROPANIZER REBOILER

Duty: 6.47 MM BTU/hr
U = 643
LMTD = 15°F
A = 670 ft^{2}F
Shell - Internal reboiler, 355°F
Tubes - 150 psig steam - condensing
Design, 150 psig, 600°F

E-15 DEPROPANIZER OVERHEAD CONDENSER

Duty: 5.37 MM BTU/hr
U = 59.3
CMTD = 21°F
A = 4,312 ft^2
Shell - Depropanizer overhead, 130°F in, 100°F out
design 336 psig, 650°F
Tubes - Cooling water 75°F in, 95°F out
design 75 psig, 650°F

FIRED HEATERS

H-1 DEPENTANIZER REBOILER

Duty: 15 MM BTU/hr at 84% efficiency (LHV) Flow: 660 GPM
(.67 sp.gr. at 430°F)
Temp: 430°F in
500°F out - 33% vapor
Pressure - 164 psig in (before control valve)
107 psig out
Design - 150 psig, 600°F

H-2 BUTANE SPLITTER REBOILER

Duty: 17.3 MM BTU/hr absorbed at 84% efficiency (LHV)
Flow: 850 GPM
Temp: 152°F in
157°F out (74% vapor)
Pressure - 170 psig in (before control valve)
120 PSIG out
Design - 165 psig, 600°F

U = Overall heat transfer rate based on outside bare tube surface BTU/HR,FT^2,°F) Include appropriate fouling factors.

CMTD = Logarithmic mean temperature difference corrected for lack of counter current flow, of 1-Shell Pass and 2-Tube Pass Exchangers.

MANUFACTURING SPECIFICATIONS

Compound (mol%)	Propane	Isobutane	N-Butane	Blending Butane
Ethane & Lighter	5 Max.			
Propane	93 Min.	3 Max.		1 Max.
Isobutane	2 Max.	92 Min.	5 Max.	
Normal Butane		7 Max.	95 Min.	
Pentane & Heavier			3 Max	20 Max.
Volatile Sulfur, GR/100 Cu. Ft.	10	10	10	10
Moisture	Pass (Cobalt Bromide)	None Free	None Free	

COST BASIS

Fuel Gas - $1.80/MM BTU/HR (LHV)

Steam, 150 psig - $2.35/1000 Lbs w/Condensate Return
50 psig - $2.00/1000 Lbs w/Condensate Return

Cooling Water - 0.08/1000 Gallons Circulated

Electric Power - 0.05/KWH

Operator Helper - $10.50/hr X 1.35 for Supervision & Benefits

Instrument Air - Negligible Incremental Cost

EQUIPMENT COSTS
(Based on Chemical Engineering Construction Index - 342.5)

Pipelines $/ft = 2.5 (Dia, IN) -2

In Plant Pipe $/ft = 2.0 (Dia, IN) +3

Pumps, $4,000 + 50 (GPM)$^{1/2}$ +3(Head, ft) 50-750 GPM 100-500 ft,HD

Motors, $55 (HP) +220 1-40 HP
$85 (HP) -1,000 50-200 HP

Motor Starters, $30 (HP) +750 Req'd for 3 HP and greater motors

Heat Exchangers 5,000 + 23 (Area, ft^2) 50-500 ft^2
10,000 + 13 (Area, ft^2) 500-2,000 ft^2

EQUIPMENT COSTS (Cont'd)

Vessels: $ = (2)(L + 2.2D)(100t + 45tD + 5D)

t = thickness, in 1/4 min.
D = diameter, ft 3-7
L = length ft.

Trays: $/Tray = 170 D Fn

D = diameter, ft 3-7
Fn = Factor for <20 Trays
Fn = 0.5 + 10/Nt (Nt=5-20)
Fn = 3 (Nt <5)

Control Systems, Installed:

$/Loop = 5600 + Cl + (1600+CF) dv

Cl = 2600 for Level Controller; else = 0
CF = 800 for Flow Controller; else = 0
dv = Diameter of Motor Valve

Pipe Insulation:

$/foot = 6 + (.35+.725i)(d+3) i = 1" -2-1/2"
$/foot = -17 + (.35+.725i)(d+14) i = 3" -4"

d = nominal diameter, in 1-10
Allow 3 feet per pipe fitting

Vessel Insulation and Fire Proofing:

$$\$ = (24+8i)(D+i/6)H + (50+14i)(D+i/6)^2 + 57\ DS$$

i = Insulation thickness
D = Vessel diameter, feet
H = Vessel height, feet
S = Skirt height, feet

E-9 SECOND STAGE COMPRESSOR AFTERCOOLER

Duty: 1.72 MM BTU/hr
U = 84
CMTD = 45.3
A = 452
Shell - hydrocarbon vapor, 180°F to 100°F
design: 233 psig, 220°F
Tubes - Cooling water, 75°F to 95°F
design - 50 psig, 650°F

E-10 N-BUTANE PRODUCT COOLER

Duty: 550,000 BTU/hr
U = 52.3
LMTD = 40°F
A = 263 ft^2
Shell - N-butane liquid, 155°F to 100°F
design - 170 psig, 400°F
Tubes - Cooling water, 75°F to 95°F
design - 50 psig, 650°F

E-11 BUTANE SPLITTER OVERHEAD CONDENSER

Duty: 16.03 MM BTU/hr
U = 125
CMTD = 27.6 A = 6,410 ft^2
Shell - Butane splitter overhead in 125°F out,
100°F design, 150 psig, 550°F
Tubes - Cooling water, 75°F to 95°F
design, 75 psig, 650°F

E-12 DEPROPANIZER FEED - BOTTOMS EXCHANGER

Duty: 2.68 MM BTU/hr
U = 84.2
CMTD = 96.9
A = 328 ft^2
Shell - Depropanizer bottoms, 355°F in, 158°F out
design, 370 psig, 650°F
Tubes - Depropanizer feed, 100°F in, 153°F out
design, 360 psig, 650°F

U = Overall heat transfer rate based on outside bare tube surface BTU/HR,FT^2,°F) Include appropriate fouling factors.

CMTD = Logarithmic mean temperature difference corrected for lack of counter current flow, of 1-Shell Pass and 2-Tube Pass Exchangers.

E-5 DEPENTANIZER FEED - BOTTOMS EXCHANGER

Duty: 3.4 MM BTU/hr
U = 53.7
CMTD = 142.2°F
A = 452 ft^2
Shell - depentanizer bottoms - 463°F to 280°F
design 120 psig, 650°F
Tubes - depentanizer feed - 186°F to 232°F
design 229 psig at 450°F

E-6 DEPENTANIZER OVERHEAD CONDENSER

Duty: 8.93 MM BTU/hr
U = 94.4
CMTD = 39.4°F
A = 2,400 ft^2
Shell - depentanizer overhead, 144°F to 100°F
design: 120 psig at 450°F
Tubes - Cooling water, 75°F to 95°F
design: 50 psig at 650°F

E-7 DEPENTANIZER LEAN OIL COOLER

Duty: 646,000 BTU/hr
U = 65
LMTD = 38
A = 263
Shell - depentanizer bottoms, 150°F to 100°F
design: 127 psig, 550°F
Tubes - cooling water, 75°F to 95°F
design: 50 psig, 650°F

E-8 FIRST STAGE COMPRESSOR AFTERCOOLER

Duty: 470,000 BTU/hr
U = 70
CMTD = 46°F
A = 147
Shell - hydrocarbon vapor, 180°F to 100°F
design: 110 psig, 65°F
Tubes - cooling water, 75°F to 95°F
design: 50 psig, 650°F

U = Overall heat transfer rate based on outside bare tube surface BTU/HR,FT^2,°F) Include appropriate fouling factors.

CMTD = Logarithmic mean temperature difference corrected for lack of counter current flow, of 1-Shell Pass and 2-Tube Pass Exchangers.

AIChE DESIGN PROBLEM
GAS CONCENTRATION UNIT
PROCESS DESCRIPTION

The Gas Concentration Unit collects various streams of gases and liquids containing low boiling saturated hydrocarbons and gases, and processes them to produce refinery fuel, propane for sale, refinery isobutane, blending butane, and pentane plus.

The refinery gases are collected into a low pressure header and compressed to 225 psia in two stages of reciprocating compressor, C-1 with inter-cooling and after-cooling to 100°F. The compressor is driven by a spark ignited internal combustion engine fueled by purchased natural gas. Instrumentation controls the speed within limits to match available gas feed, with fuel gas make-up to prevent vacuum on the compressor suction when the engine is at minimum speed.

The refinery liquids are collected into a high pressure header and flow into the de-ethanizer, V-1. The purchased NGL arrives by high pressure pipeline, is flow controlled into the de-ethanizer. Naphtha (a liquid characterized as mixture of nonane, decane, and undecane) is produced in the crude unit, is used as lean oil to absorb the propane preventing excessive loss to refinery fuel. The use of refrigeration would cause freezing due to possible water content of feeds. Depentanized oil is used to supplement the naphtha to maintain 75 GPM flow to the deethanizer. A series of 3 pumparound coolers prevent excessive temperatures in the top of the absorber. The internal tray take-off arrangement allows the pump flow to be set constant with excess cooled liquid returning back to pump suction after blending with liquids from the tray above. The absorber is simulated as having 16 ideal stages, numbered from the bottom with liquid cooling to 100°F on Stages 15, 14 and 13; feed enters on Stage 10; the reboiler is Stage 1.

If water enters in excess of saturated limit in hydrocarbons, provision is made to collect the water in V-15 and be drained manually. Bottom tray liquid flows to the reboiler (E-4) by gravity and bottom product is level controlled to the depentanizer, V-2. Hot liquid flow through the reboiler is controlled as necessary to control the ethane and H_2S content at a low level.

The de-ethanizer bottoms stream enters the depentanizer (V-2) through a bottoms-feed exchanger (E-5) into tray 14 of 30 valve trays. The depentanizer is operated as a partial depentanizer to allow as much normal pentane as possible to leave with the lean oil and be up-graded with the naphtha in the isomerization unit. The depentanized oil is cooled by air fin cooler (A-1) and trim cooled against cooling water in exchanger E-7. The DeC_5 reboiler circulation, pumped by P-4, provides heat as necessary to E-4 and enters the convection section of the heater (H-1), then passes through the fired section and emerges as a mixture of vapor and liquid returning to the bottom of V-2 for separation. The fired heater fuel rate is controlled to hold the outlet temperature and outlet temperature is changed as necessary to control the fractionator.

The overhead vapors are condensed against cooling water in E-6 and enter the overhead accumulator (V-3). A small vapor stream is by-passed to maintain column pressure. If water is present it can be drained from the water leg manually. The reflux is pumped and flow controlled to the Tray 30. The depentanizer is simulated using 22 stages with feed on stage 10 from bottom. The NGL is pumped using P-11 and level controlled to the treaters.

A-1 LEAN OIL AIR FIN COOLER

Duty: 2.36 MM BTU/hr
U = 89 (overall, bare)
Oil Side: 290°F in, 88 psig
150°F out, 78 psig
Air Side: 100°F in
145°F out
Area: 297 ft^2 bare, 4,772 ft^2 finned.

C-1 GAS COMPRESSOR

Reciprocating Package Unit to compress 2.4 MM SCFD from 15.6 psia to 225 psia. Elevation is 4,260' above sea level or 12.6 psia atm. pressure. Compressor has a speed controller to adjust engine speed to compensate for varying suction volume.

HEAT EXCHANGERS

E-1, E-2, E-3 DE-ETHANIZER INTERCOOLERS

Each: Double pipe multi-tube
6-5/8" O.D. X 15'
Duty: 795,000 BTU/hr
U = 55.8
LMTD = 27.4
Area = 520 ft^2 (bare tube)
Water in tubes: 75°F in, 95°F out
Hydrocarbon in shell: 125°F in, 100°F out
Design: shell and tubes 500 psig, 650°F.

E-4 DE-ETHANIZER REBOILER

Duty: 6.55 MM BTU/hr
U = 66.1
CMTD = 216°F
A = 459 ft^2
Depentanizer bottoms in tube - 463°F in, 398°F out
De-ethanizer tower bottoms in shell - 184°F in, 213°F out
41.4% vaporized
Design: Shell - 275 psig, 235°F
Tubes - 270 psig, 650°F.

U = Overall heat transfer rate based on outside bare tube surface BTU/HR,FT^2,°F) Include appropriate fouling factors.

CMTD = Logarithmic mean temperature difference corrected for lack of counter current flow, of 1-Shell Pass and 2-Tube Pass Exchangers.

The butane splitter (V-13) had been used in another service previously. The feed enters on tray 28 of 50. The vapors are condensed against cooling water in E-11 and accumulated in V-14. Overhead liquids are pumped with P-9; reflux is flow controlled and the isobutane is level controlled to storage. The column is simulated using 42 stages. The reboil vapors are generated by reboiler recycle pumping and flow controlled through a fired heater (H-2). The blending butane is cooled against cooling water using E-10 and level controlled to storage.

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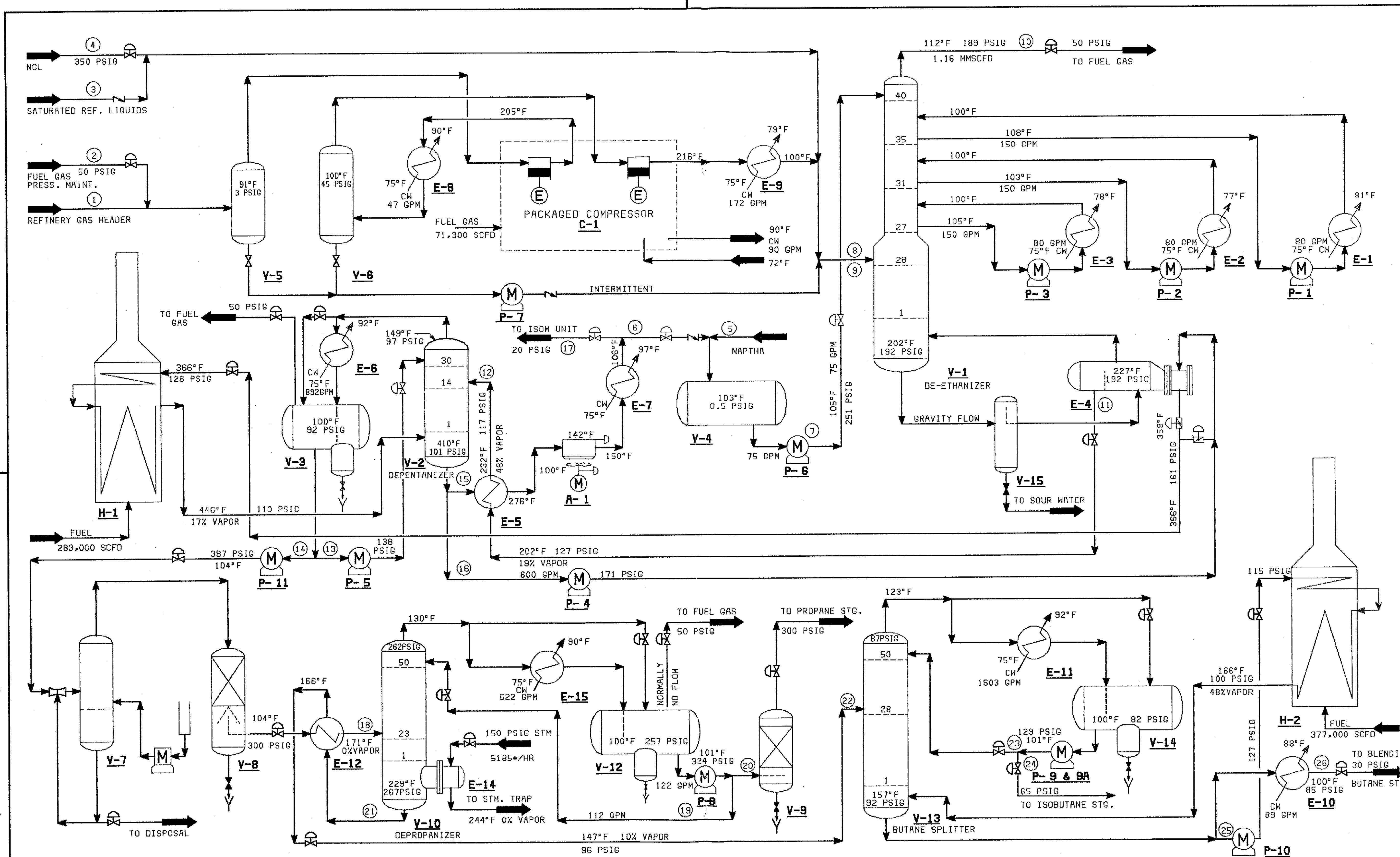
V-7 is an aqueous liquid caustic treater. The feed induces a flow of caustic through an eductor where intensive mixing occurs. Any H_2S or CO_2 contained in the NGL will react with caustic to form the sodium salt and be removed. Separation of the two liquid phases occurs in vessel V-7 with final separation of small droplets in V-8. A small reciprocating injection pump is used to add aqueous sodium hydroxide to maintain caustic concentration and the excess is removed by interface level control to disposal.

The depropanizer (V-10) had been used in another service previously. The treated NGL passes through a back pressure control valve and then through a bottoms-feed exchanger (E-12), and enters the depropanizer (V-10) on tray 23 of 50 valve trays. The overhead vapors are condensed against cooling water at E-15 and accumulated in V-12. The product and reflux are pumped by P-8. The column pressure is controlled by a condenser by-pass. Reflux rate is flow controlled and the propane product is level controlled through V-9 to storage. The reboiler is an internal bundle condensing steam to reboil heat. Steam flow is controlled to control the concentration of propane in the bottoms. The butanes plus are level controlled through E-12 to the butane splitter.

V-9 is a dehydrator composed of solid balls of sodium hydroxide about 3/8" diameter. Water dissolved in the propane contacts the solid caustic forming an aqueous solution which flows to the bottom of the vessel where it is manually drained. V-9 is reloaded with caustic balls as required, designed to last about one year.



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FOR BIDS
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PROCESS FLOW SHEET
GAS CONCENTRATION UNIT
OPERATIONS, 1990

JA NO.
FILE CODE
AFE NO.
SCALE
DWG NO.
SH NO. –

CAD# OBPCLS