Korea Process Simulation Olympiad 2016
Problem Statement

Disclaimer

All the data on this problem statement are hypothetical values and/or equations merely for this Olympiad which is dedicated to the education of University students. The data can NOT be applied to any other usage regardless to commercial or non-commercial purposes. The data is subject to change even at subsequent Olympiads. Olympiad committee or sponsors NEVER take responsibility for damage or inconvenience caused by using the data.

Title: Feasibility study of gas processing plant

1. Background:

Gas processing plant produces a variety of products like NGL, LNG, LPG and natural gasoline and usually consists of gathering system, field operation, gas treating and dehydration, hydrocarbon recovery and liquefaction. The configuration of the plant highly depends on the composition and feed gas, required product by market, size of the well and location.

2. Problem:

Assume you are working for Global Investment Company. Your company has succeed in finding natural gas well in East Timor but the size is smaller than originally expected so want to do feasibility study for building a gas plant there.
The market needs sales gas, LPG and condensate. The local government ask your company to add CO2 capture facility for environmental issues
So the role of your engineering team is doing feasibility study with conceptual design and economic analysis based on the requirements.

2-1.Capacity: 70 MMSCFD of natural gas processing for 20 years

2-2.Process configuration: Gas gathering station, Slug catcher, dehydration (dryer with molecular sieve), carbon capture and compression, deethanizer and debutanzier

2-3.Gas specification: Gas composition at slug catcher pressure control valve outlet and water is saturated.

Component	mol %
Methane	69.8
Ethane	9.3
Propane	7.3
i-Butane	1.6
n-Butane	2.7
i-Pentane	1.4
n-Pentane	1.7
C6	0.7
C7	0.2
Nitrogen	0.2
CO_2	5.1
Total Sulfur	NIL
	100.00

Pressure : 5 kgf/cm2
Temperature : 50 degC

2-4.Product specification & BL condition:

Sales gas purity ; Predominantly C1 & C2
C3 and heavier Max. 3 mol%
CO_2 Max. 100 ppmv

LPG ; Predominantly C3 & C4 ;
TVP Max. 208 Psig at 37.8degC
C5+ Heavier Max. 0.1 wt%

Condensate : C5 & heavier; Max. 1.0 liquid vol% of C4

BL (battery limit) condition for sales gas :
Temperature : Max. 49 deg C
Pressure: 50 kgf/cm2

2-5.CO_2 capture and compression

CO_2 capture solvent; 20wt% MEA (Ethanolamine) solution
Lean/rich amine loading (CO_2 mol/ MEA mol) : max. 0.1 / max. 0.4
CO_2 in treated gas (from absorber) : max. 50ppmw
CO_2 storage facility condition : 35 degC, 85 kgf/cm2
CO_2 product spec ; H_2O max.1000ppmv / purity min.99mol%

2-6.Utility:

CW	32degC / 42 degC
High pressure steam	40 kg/cm^{2}g
Low pressure steam	4 kg/cm^{2}g

2-7.Economic conditions

Sales gas = US $ 5/MMBTU-
LPG = US $ 790/TON
Condensate = US $ 50/Barrel

Electricity: 0.08 US $/KWH
Steam/Fuel cost: US $65/$10^6$ kcal
Operation time: 8000 h/year

Cost of each unit is roughly estimated by the procedure listed in APPENDIX-1 using Marshal & Swift Equipment Cost Index (M&S). We assume M&S =1600.
Cooling water cost: 0.03 US$/m^3

2-8.Physical property estimation:

Process side: applicant need to select
Utility steam: steam table IPWS IF97

3. Report

Applicants need to submit the final report which contains simulation validation report with schematic drawing and material balance, equipment list and economic analysis.

3-1. Simulator validation report

You need to convince your executive of your company that your simulation results are right. To do so, you need to submit a simulator validation report which should contain selected process configuration, the physical properties and estimation methods applied to the work.

- Pure component parameters and thermodynamics used in the simulation
- Reason of the selection of the method must be stated.
- Selection of process configuration

3-2.Process flow diagram (PFD) with material/energy balances

Snapshots of flow sheet of major process simulators are acceptable. Temperature, pressure, flow rate and composition of each stream must be indicated on the PFD. Heating or cooling duty of each equipment should be indicated also.

If, the simulator does not produce a flow diagram, applicants can draw it on major drawing software like Microsoft VISIO, Excel or AutoCAD.

Unit of measure should be metric.

3-3. Economic analysis

Estimate total investment cost and annual operating cost based on given information and your assumptions.

4. Evaluation (scoring) guideline

4-1. Screening by important requirements

Reporting all the above mentioned.
Selection of correct thermo methods.
Usage of correct way of indicating PFD.

4-2. Scoring by the economic analysis

Reporting all the reason and assumption of estimated commercial value. Well estimation of TIC and operating cost.

4-3. Scoring by process configuration

Good safety consideration
Cost conscious
Operability

4-4. Scoring by simulation skill

Effective usage of simulator tools

APPENDIX-1 Estimation Procedure for Installation Cost

Although cost estimation is not due, if applicant wants to roughly estimate the installation cost, use the following equations which are indicated by James M.Douglas “Conceptual Design of Chemical Processes” , McGRAW-HILL International editions (1988).

We assume M&S =1600.

Fixed cost

1. TIC(Total Investment Cost) for gathering center, slug catcher and dehydration unit is 0.1 billion USD.

The equipment cost for ‘dryer regeneration gas heater’ and ‘regeneration gas cooler’ shall be estimated and included.

2. Process condition for dryer regeneration gas heater

Fluid Allocation		Hot Fluid		Cold Fluid	
Fluid Name		HP Steam		Regeneration Gas	
Fluid Quantity, Total	KG/HR			12,995	
Vapor (In/Out)	KG/HR			12,995	12,995
Liquid (In/Out)	KG/HR				
Temperature (In/Out)	°C			72.3	232.0
Density (Vapor/Liquid)	KG/CUM	/	/	1.87 /	1.21 /
Viscosity (Vapor/Liquid)	CP	/	/	0.01 /	0.02 /
Molecular Weight (Vapor/Liquid)		/	/	8.36 /	8.36 /
Specific Heat (Vapor/Liquid)	KJ/KG K	/	/	3.972 /	4.444 /
Thermal Conductivity (Vapor/Liquid)	W/ M K	/	/	0.104 /	0.149 /
Latent Heat					
Inlet Pressure	Kgf/cm2(A)			6.50	
Pressure Drop, Allowable	Kgf/cm2(A)	0.7		0.70	
Heat Exchanged	MW			2.421	
Design Margin on Flows & Duty		10%			

3. Process condition for dryer regeneration cooler

Fluid Allocation		Hot Fluid		Cold Fluid	
Fluid Name		Wet Regeneration Gas		Cooling Water	
Fluid Quantity, Total	KG/HR	14,372			
Vapor (In/Out)	KG/HR	14,372	13,561		
Liquid (In/Out)	KG/HR		811		
Temperature (In/Out)	°C	150.0	45.0		
Density (Vapor/Liquid)	KG/CUM	1.31 /	1.58 / 990.45	/	/
Viscosity (Vapor/Liquid)	CP	0.02 /	0.01 / 0.61	/	/
Molecular Weight (Vapor/Liquid)		8.81 /	8.55 / 18.02	/	/
Specific Heat (Vapor/Liquid)	KJ/KG K	3.970 /	3.821 / 4.121	/	/
Thermal Conductivity (Vapor/Liquid)	W/ M K	0.120 /	0.094 / 0.651	/	/
Critical Pressure (Liquid)	Kgf/cm2(A)		22,118.0		
Latent Heat					
Inlet Pressure	Kgf/cm2(A)	5.20			
Pressure Drop, Allowable	Kgf/cm2(A)	0.70		0.7	
Heat Exchanged	MW	2.144			
Design Margin on Flows & Duty		30%			

Heat Exchangers

Mid-1968 cost, shell and tube, complete fabrication.

$$\text{Purchased Cost, \$} = \left(\frac{\text{M\&S}}{280}\right)(101.3A^{0.65}F_c)$$

where A = area ft^2; $200 < A < 5000$

$$F_c = (F_d + F_p)F_m$$

Shell-and-Tube Material = F_m

Surface area, ft^2	CS/ CS	CS/ Brass	CS/ MO	CS/ SS	SS/ SS	CS/ Monel	Monel/ Monel	CS/ T_i	T_i/ T_i
1000 to 5000	1.00	1.30	2.15	2.81	3.75	3.10	4.25	8.95	13.05

$$\text{Installed Cost, \$} = \left(\frac{\text{M\&S}}{280}\right)101.3A^{0.65}(2.29 + F_c)$$

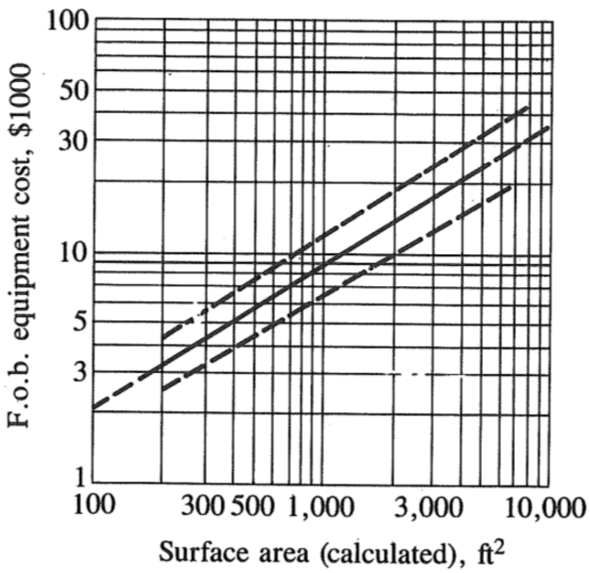


FIGURE E.2-3
Shell-and-tube heat exchangers. [*K. M. Guthrie, Chem. Eng.*, **76**(6): *114* (*March 24, 1969*).]

TABLE E.2-3
Correction factors for heat exchangers

Design type	F_d	Design pressure, psi	F_p
Kettle, reboiler	1.35	Up to 150	0.00
Floating head	1.00	300	0.10
U-tube	0.85	400	0.25
Fixed-tube sheet	0.80	800	0.52
		1000	0.55

Spiral Plate Heat Exchanger
1st Quarter 1998 dollars

Heat Transfer Area, (Square feet)	Purchased Equipment Cost ($)	Installed Cost ($)
40	$6,700	$19,200
100	$9,100	$25,100
200	$13,200	$34,000
300	$21,100	$49,400
400	$25,500	$57,400
500	$29,900	$65,000
600	$34,400	$72,400
700	$42,600	$85,300
800	$35,500	$74,200
900	$40,000	$81,300
1,000	$44,700	$88,500
1,100	$49,600	$95,700
1,200	$54,700	$102,900
1,300	$60,100	$110,400

The overall heat transfer coefficient for plate type heat exchanger / spiral heat exchanger is assumed 500 Btu/(ft2 F h)

The cost for plate type heat exchanger is equivalent to spiral plate heat exchanger.

Gas Compressors

Mid-1968 cost, centrifugal machine, motor drive, base plate and coupling.

$$\text{Purchased Cost, \$} = \left(\frac{\text{M\&S}}{280}\right)(517.5)(\text{bhp})^{0.82}F_c$$

where bhp = brake horsepower; 30 < bhp < 10,000

$$F_c = F_d$$

$$\text{Installed Cost, \$} = \left(\frac{\text{M\&S}}{280}\right)(517.5)(\text{bhp})^{0.82}(2.11 + F_c)$$

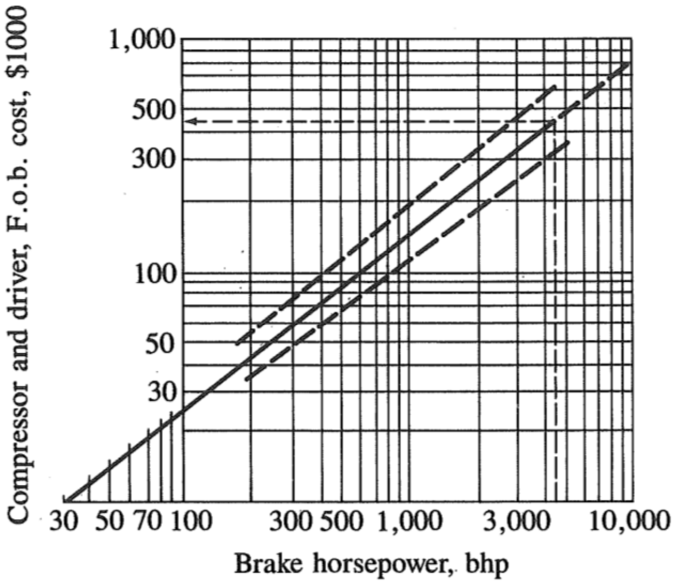


FIGURE E.2-4
Process gas compressors and drives. [*K. M. Guthrie, Chem. Eng.*, **76**(*6*): *114* (*March 24, 1969*).]

TABLE E.2-4
Correction factors for Compressors

Design type F_d	Factor
Centrifugal, motor	1.00
Reciprocating, steam	1.07
Centrifugal, turbine	1.15
Reciprocating, motor	1.29
Reciprocating, gas engine	1.82

Gas Turbine
1st Quarter 1998 dollars

Power Output (Horsepower)	Purchased Equipment Cost ($)	Installed Cost ($)
1,000	$476,200	$565,200
5,000	$1,254,100	$1,376,400
10,000	$1,903,000	$2,051,300
50,000	$9,639,300	$9,975,400
100,000	$16,148,100	$16,738,600
150,000	$21,837,300	$22,659,400
200,000	$27,052,000	$28,056,000
250,000	$31,940,100	$33,192,400
300,000	$36,583,000	$37,998,000
350,000	$41,031,000	$42,609,000
370,000	$42,764,000	$44,407,000

Steam Turbine
1st Quarter 1998 dollars

Power Output (Horsepower)	Purchased Equipment Cost ($)	Installed Cost ($)
10	$19,100	$36,000
50	$25,200	$46,500
100	$28,500	$53,600
500	$37,700	$108,800
950	$42,100	$126,700
1,000	$85,000	$169,800
2,500	$269,000	$364,400
5,000	$575,000	$688,000
7,500	$781,400	$907,900
10,000	$971,400	$1,106,600
15,000	$1,320,100	$1,477,100
20,000	$1,641,100	$1,825,200
30,000	$2,230,200	$2,447,300

Expander
The cost of expender is 50% of steam turbine at same power.

Pressure Vessels, Columns, Reactors

$$\text{Purchased Cost, \$} = \left(\frac{M\&S}{280}\right)(101.9D^{1.066}H^{0.82}F_c)$$

where D = diameter, ft

H = height, ft

$F_c = F_m F_p$

Pressure	Up to 50	100	200	300	400	500	600	700	800	900	1000
F_p	1.00	1.05	1.15	1.20	1.35	1.45	1.60	1.80	1.90	2.30	2.50

$$\text{Installed Cost, \$} = \left(\frac{M\&S}{280}\right)101.9D^{1.066}H^{0.802}(2.18 + F_c)$$

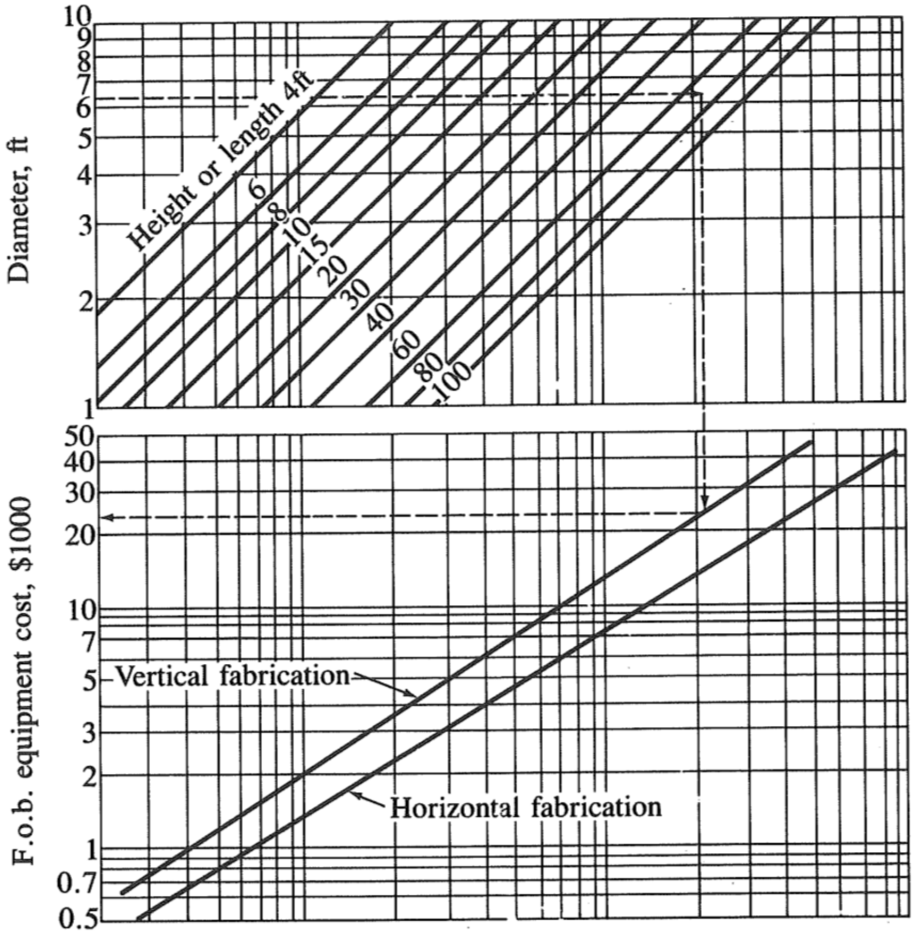


FIGURE E.2-5
Pressure vessels. [*K. M. Guthrie, Chem. Eng.*, **76**(*6*): *114* (*March 24, 1969*).]

TABLE E.2-5
Correction factors for pressure vessels

Shell material	CS	SS	Monel	Titanium
F_m, clad	1.00	2.25	3.89	4.25
F_m, solid	1.00	3.67	6.34	7.89

Distillation Column Trays and Tower Internals

$$\text{Installed Cost, \$} = \left(\frac{\text{M\&S}}{280}\right) 4.7 D^{1.55} H F_c$$

where D = diameter, ft

H = tray stack height, ft (24-in. spacing)

$F_c = F_s + F_t + F_m$

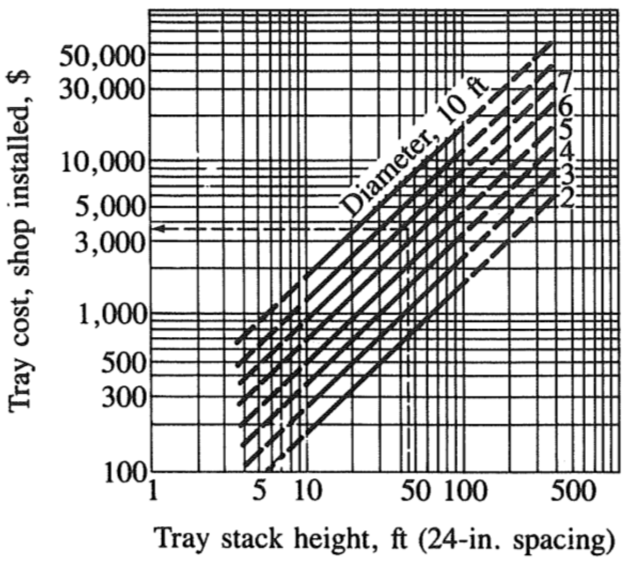


FIGURE E.2-6
Distillation column trays. [*K. M. Guthrie, Chem. Eng.*, **76**(6): *114* (*March 24, 1969*).]

TABLE E.2-6
Correction factors for column trays

Tray spacing, in.	24	18	12				
F_s	1.0	1.4	2.2				
Tray type		Grid (no down-comer)	Plate	Sieve	Trough or valve	Bubble cap	Koch Kascade
F_t	0.0		0.0	0.0	0.4	1.8	3.9
Tray material	CS	SS	Monel				
F_m	0.0	1.7	8.9				

TABLE E.2-7
Tower packings

Material	Materials and labor, $/ft^3$
Activated carbon	14.2
Alumina	12.6
Coke	3.5
Crushed limestone	5.8
Silica gel	27.2
1-in. Raschig rings—Stoneware	5.2
Porcelain	7.0
Stainless	70.2
1-in. Berl saddles—Stoneware	14.5
Porcelain	15.9

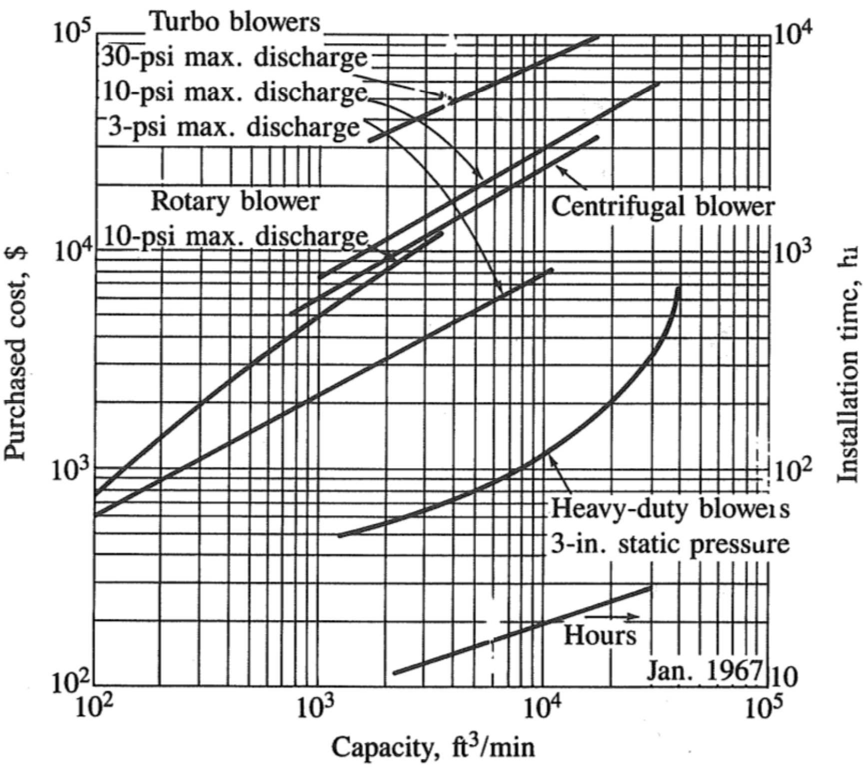
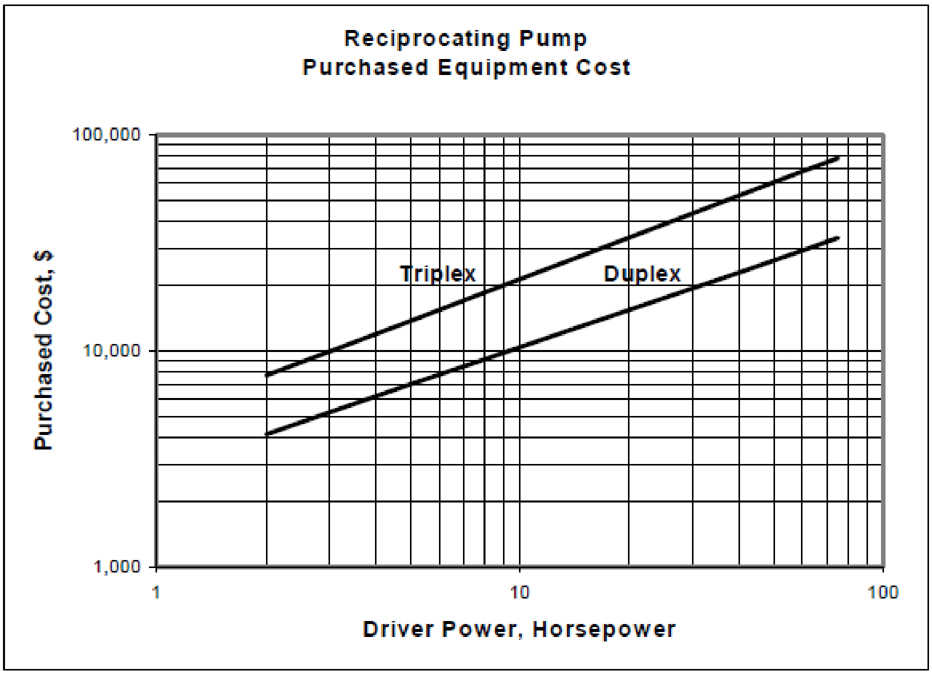


FIGURE E.2-7
Blowers (heavy-duty, industrial type). (*From M. S. Peters and K. D. Timmerhaus, Plant Design and Economics for Chemical Engineers, 3d ed., McGraw-Hill, New York, 1980, p. 562.*)

Reciprocating Pump

Description: Reciprocating duplex with steam driver. Triplex (plunger) with pump-motor driver.

Design Basis:
1st Quarter 1998 Dollars
Material: Carbon Steel
Design Temperature: 68 °F
Liquid Specific Gravity: 1
Efficiency: 82%



Reciprocating Pump
1st Quarter 1998 dollars

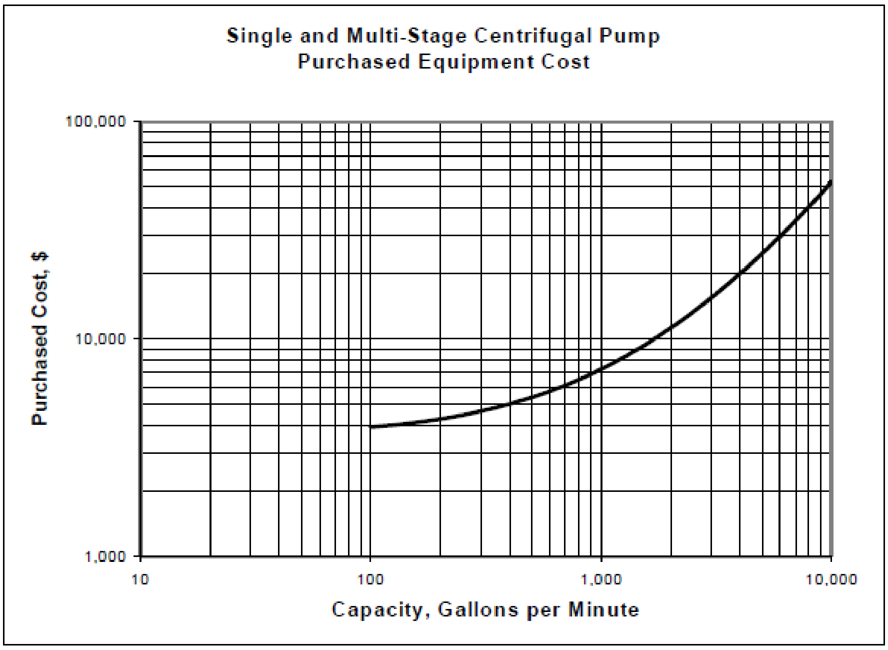
Capacity (Gallons/ minute)	Driver Power (Horse-power)	Duplex Purchased Equipment Cost ($)	Installed Cost ($)	Triplex Purchased Equipment Cost ($)	Installed Cost ($)
25	2	$4,100	$10,600	$7,700	$15,500
50	5	$7,000	$14,600	$13,800	$22,700
100	7.5	$8,800	$17,800	$17,900	$28,200
200	15	$13,100	$22,500	$27,900	$38,600
300	25	$17,600	$28,800	$38,700	$51,200
400	30	$19,600	$31,000	$43,500	$56,200
500	40	$23,100	$34,700	$52,300	$65,300
600	50	$26,300	$38,100	$60,300	$73,400
700	60	$29,200	$43,700	$67,800	$83,700
800	60	$29,200	$43,700	$67,800	$83,800
900	75	$33,300	$48,100	$78,200	$94,500
1,000	75	$33,300	$48,200	$78,200	$94,500

Centrifugal Pump

Description: Single and multistage centrifugal pumps for process or general service when flow/head conditions exceed general service. Split casing not a cartridge or barrel. Includes standard motor driver.

Design Basis:
1st Quarter 1998 Dollars

Material:	Carbon Steel
Design Temperature:	120 °F
Design Pressure:	150 psig
Liquid Specific Gravity:	1
Efficiency:	<50 GPM = 60%
	50 – 199 GPM = 65%
	100 – 500 GPM = 75%
	> 500 GPM = 82%
Driver Type:	Standard motor
Seal Type:	Single mechanical seal



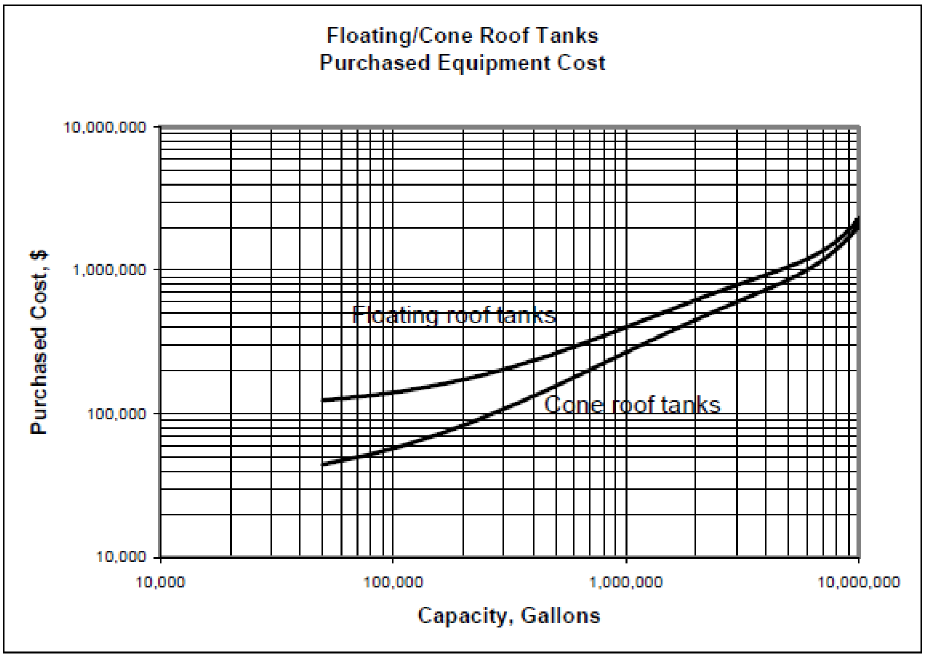
Centrifugal Pump
1st Quarter 1998 dollars

Capacity (Gallons/ minute)	Purchased Equipment Cost ($)	Installed Cost ($)
100	$3,400	$22,800
200	$4,100	$23,800
300	$4,700	$27,700
400	$5,300	$28,500
500	$5,800	$29,000
1,000	$8,700	$37,500
2,000	$10,200	$44,800
3,000	$15,200	$58,100
4,000	$19,500	$72,300
5,000	$23,800	$77,100
6,000	$28,400	$93,400
7,000	$37,800	$103,000
8,000	$41,300	$119,700
9,000	$47,300	$126,200
10,000	$51,200	$144,800

Storage Tanks

Description:
Floating Roof: Typically constructed from polyurethane foam blocks or nylon cloth impregnated with rubber or plastic, floating roofs are designed to completely contact the surface of the storage products and thereby eliminate the vapor space between the product level and the fixed roof. Floating roof tanks are suitable for storage of products having vapor pressure from 2 to 15 psia.
Cone Roof: Typically field fabricated out of carbon steel. They are used for storage of low vapor pressure (less than 2 psia) products, typically ranging from 50,000 – 1,000,000 gallons.

Design Basis:
1st Quarter 1998 Dollars
Shell Material: A515
(Carbon Steel Plates for pressure vessels for intermediate and higher temperature service)
Design Temperature: 650 °F
Design Pressure: 15 psig
Diameter: 2 – 14 feet
Length: 4.3 – 81 feet
Total Weight: 1100 –59,400 pounds



Storage Tanks
1st Quarter 1998 dollars

Diameter (Feet)	Height (Feet)	Total Weight (Pounds)	Capacity (Gallons)	Purchased Equipment Cost ($)	Installed Cost ($)
Floating Roof					
17.0	32.0	41,300	50,000	$118,000	$163,400
20.0	32.0	46,700	75,000	$128,200	$180,700
24.0	32.0	55,000	100,000	$143,200	$205,100
37.0	32.0	89,300	250,000	$197,700	$250,000
47.0	40.0	142,400	500,000	$267,800	$332,400
57.0	40.0	195,000	750,000	$335,700	$411,700
66.0	40.0	245,700	1,000,000	$396,600	$480,200
134.0	48.0	858,900	5,000,000	$1,061,200	$1,250,900
175.0	56.0	2,219,100	10,000,000	$2,273,000	$2,564,300
Cone Roof					
17.0	32.0	21,000	50,000	$42,400	$87,800
20.0	32.0	26,400	75,000	$48,900	$101,400
24.0	32.0	34,800	100,000	$59,200	$121,100
37.0	32.0	69,400	250,000	$98,600	$150,900
47.0	40.0	123,100	500,000	$157,800	$222,400
57.0	40.0	176,400	750,000	$214,800	$296,800
66.0	40.0	228,000	1,000,000	$266,100	$349,700
134.0	48.0	853,600	5,000,000	$864,300	$1,054,000
175.0	56.0	2,226,100	10,000,000	$2,040,700	$2,332,000

Refrigeration: $C = 178FQ^{0.65}$ K\$, $0.5 < Q < 400$ M Btu/hr, installed prices

Temperature Level (°C)	*F*
0	1.00
–10	1.55
–20	2.10
–30	2.65
–40	3.20
–50	4.00

Equipment cost for Refrigeration system is assumed two times of installation cost.