Solutions Manual for Analysis, Synthesis, and Design of Chemical Processes

Third Edition

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Chapter 1

1.1 Block Flow Diagram (BFD)
Process Flow Diagram (PFD)
Piping and Instrument Diagrams (P&ID)

(a) PFD
(b) BFD
(c) PFD or P&ID
(d) P&ID
(e) P&ID

1.2 P&ID

1.3 It is important for a process engineer to be able to review a 3-dimensional model prior to the construction phase to check for clearance, accessibility, and layout of equipment, piping, and instrumentation.

1.4 (1) Clearance for tube bundle removal on a heat exchanger.
(2) NPSH on a pump – affects the vertical separation of feed vessel and pump inlet.
(3) Accessibility of an instrument for an operator – must be able to read a PI or change/move a valve.
(4) Separation between equipment for safety reasons – reactors and compressors.
(5) Crane access for removing equipment.
(6) Vertical positioning of equipment to allow for gravity flow of liquid.
(7) Hydrostatic head for thermosiphon reboiler – affects height of column skirt.

1.5 Plastic models are no longer made because they are too expensive and difficult to change/revise. These models have been replaced with virtual/E-model using 3-D CAD. Both types of model allow revision of critical equipment and instrument placement to ensure access, operability, and safety.

1.6 Another reason to elevate the bottom of a tower is to provide enough hydrostatic head driving force to operate a thermosiphon reboiler.

1.7 (a) PFD or P&ID
(b) PFD
(c) PFD
(d) P&ID
(e) BFD (or all PFDs)

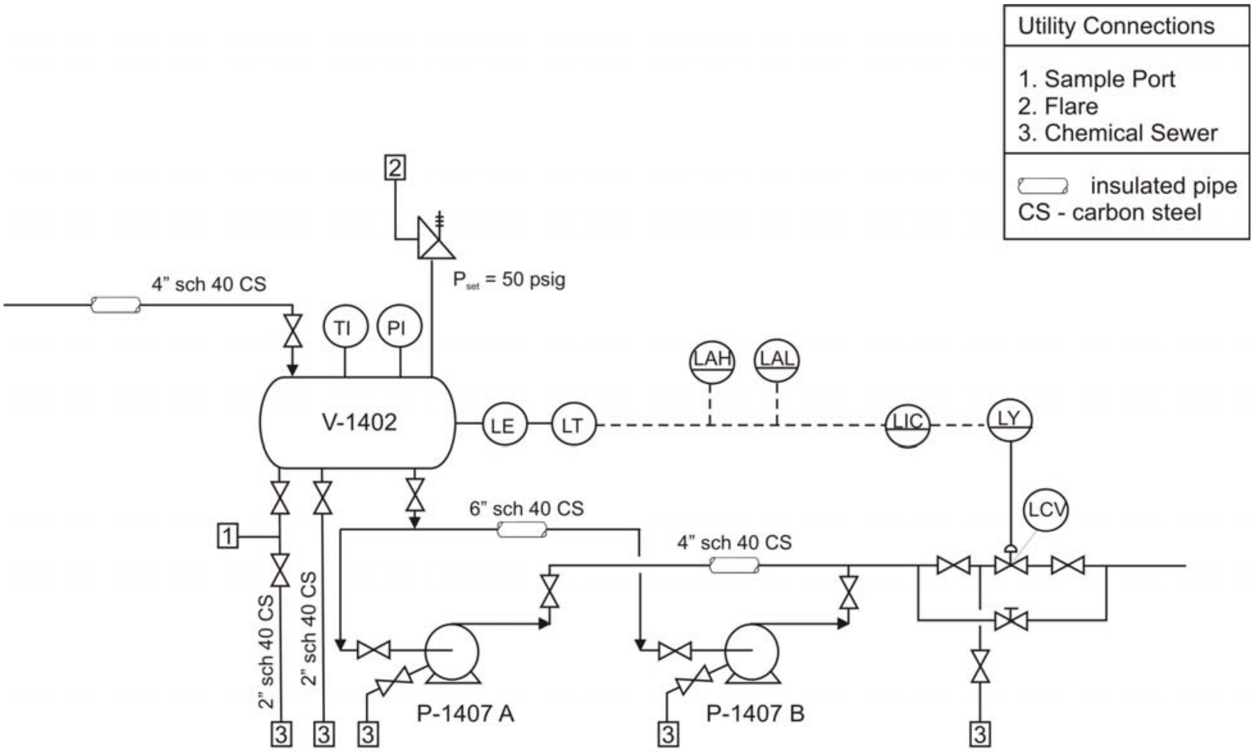
1.8 A pipe rack provides a clear path for piping within and between processes. It keeps piping off the ground to eliminate tripping hazards and elevates it above roads to allow vehicle access.

1.9 A structure – mounted vertical plant layout is preferred when land is at a premium and the process must have a small foot print. The disadvantage is that it is more costly because of the additional structural steel.

1.10 (a) BFD – No change
PFD – Efficiency changed on fired heater, resize any heat exchanger used to extract heat from the flue gas (economizer)
P&ID – Resize fuel and combustion air lines and instrumentation for utilities to fired heater. Changes for design changed of economizer (if present)
(b) BFD – Change flow of waste stream in overall material balance
PFD – Change stream table
P&ID – Change pipe size and any instrumentation for this process line
(c) BFD – No change
PFD – Add a spare drive, e.g. D-301 → D-301 A/B
P&ID – Add parallel drive
(d) BFD – No change
PFD – No change
P&ID – Note changes of valves on diagram

1.11 (a) A new vessel number need not be used, but it would be good practice to add a letter to donate a new vessel, e.g. V-203 → V-203N. This will enable an engineer to locate the new process vessel sheet and vendor information.
(b) P&ID definitely
PFD change/add the identifying letter.

1.12



Solution to Chapter 1 - Problem 12

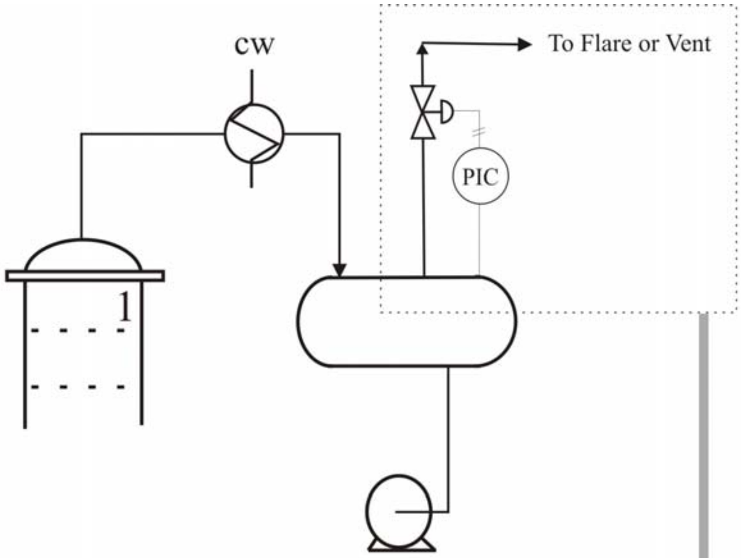
1.13 (a)
- (i) Open globe valve D
- (ii) Shut off gate valves A and C
- (iii) Open gate valve E and drain contents of isolated line to sewer
- (iv) Perform necessary maintenance on control valve B
- (v) Reconnect control valve B and close gate valve E
- (vi) Open gate valves A and C
- (vii) Close globe valve D

(b) Drain from valve E can go to regular or oily water sewer.

(c) Replacing valve D with a gate valve would not be a good idea because we loose the ability to control the flow of process fluid during the maintenance operation.

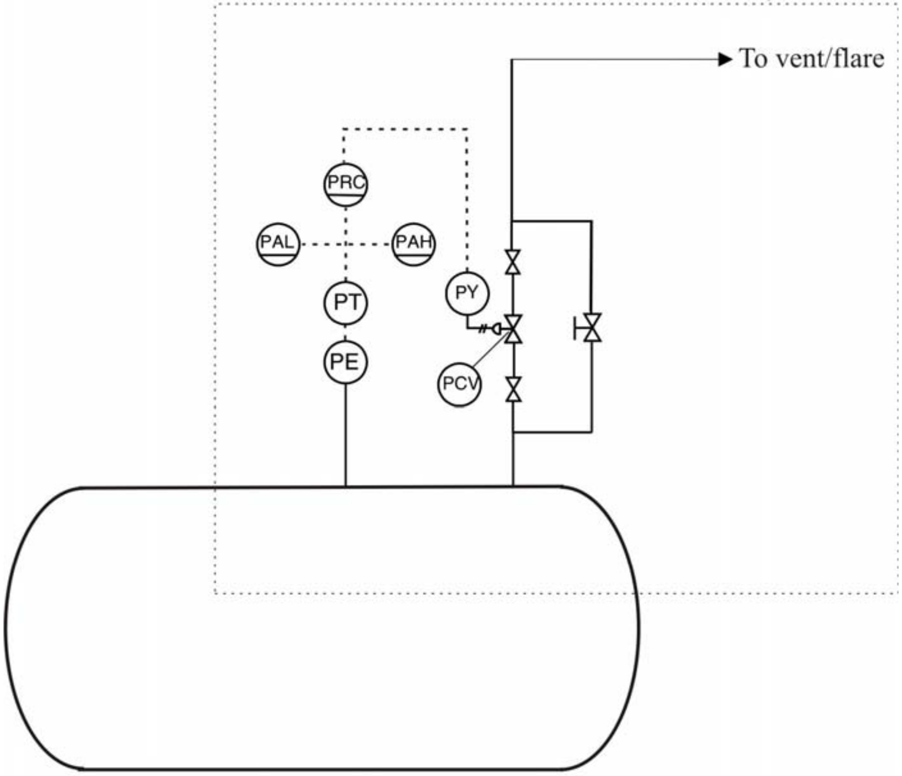
(d) If valve D is eliminated then the process must be shut down every time maintenance is required on the control valve.

1.14



Solution to Chapter 1 - Problem 14

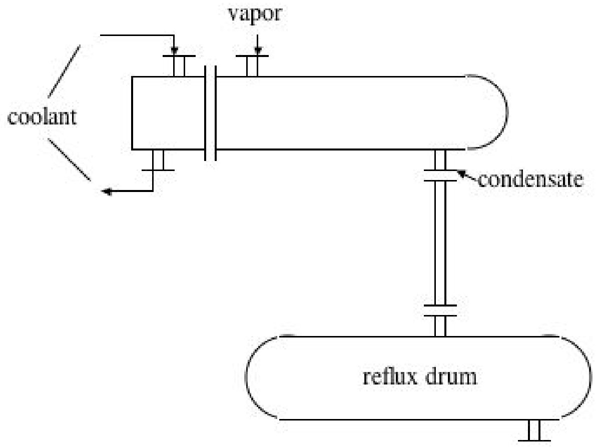
1.15



Solution to Chapter 1 - Problem 15

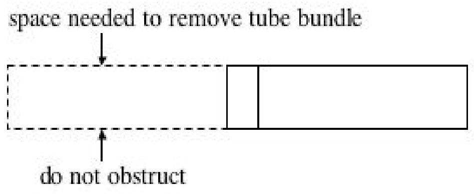
1.16 (a) For a pump with a large NPSH – the vertical distance between the feed vessel and the pump inlet must be large in order to provide the static head required to avoid cavitating the pump.

b) Place the overhead condenser vertically above the reflux drum – the bottom shell outlet on the condenser should feed directly into the vertical drum.



c) Pumps and control valves should always be placed either at ground level (always for pumps) or near a platform (sometimes control valves) to allow access for maintenance.

d) Arrange shell and tube exchangers so that no other equipment or structural steel impedes the removal of the bundle.



e) This is why we have pipe racks – never have pipe runs on the ground. Always elevate pipes and place on rack.

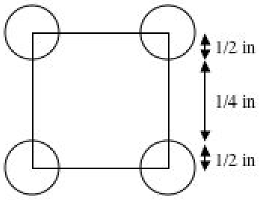
f) Locate plant to the east of major communities.

1.17

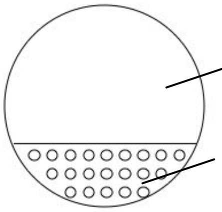
1.17 HT area of 1 tube = $\pi DL = \pi\left(\frac{1}{12}\right)(12\text{ ft}) = 3.142\text{ ft}^2$

Number of tubes = $(145\text{ m}^2)\cdot\left(\frac{3.2808\text{ ft}}{\text{m}}\right)^2\left(\frac{1}{3.142\text{ ft}^2}\right) = 497\text{ tubes}$

Use a 1 1/4 inch square pitch ⇒



Fractional area of the tubes = $\frac{\pi}{4}\left(\frac{1\text{ m}}{1.25\text{ in}}\right)^2 = 0.5027\left(\frac{\text{m}}{\text{in}}\right)^2$



$A_{VAP} = 3\ A_{LIQ} \therefore CSA_{SHELL} = 4\ A_{LIQ}$

$A_{LIQ} = \left(\frac{497}{0.5027}\right)\left(\frac{\text{in}}{\text{m}}\right)^2\left(\frac{\pi}{4}\right)(1\text{ m})^2 = 777\text{ in}^2$

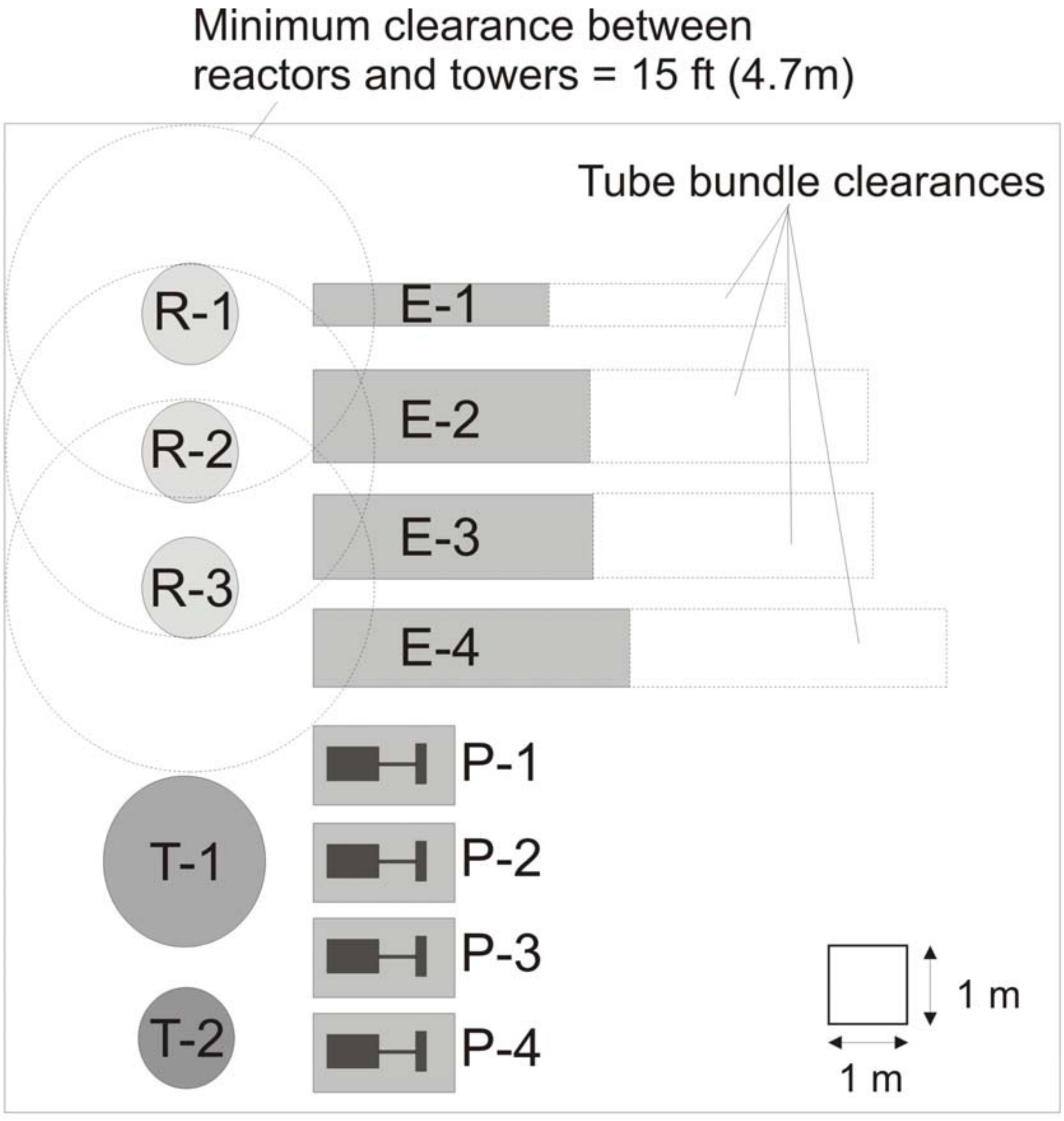
$CSA_{SHELL} = (4)(777) = 3108\text{ in}^2 \Rightarrow \frac{\pi}{4}D^2{}_{SHELL} = 3108\text{ in}^2$

$D_{SHELL} = \sqrt{\frac{(4)(3108\text{ in}^2)}{\pi}} = 62.9\text{ in} = 1.598\text{ m}$

Length of Heat Exchanger = $(2+12+2)\text{ ft} = 16\text{ ft} = 4.877\text{ m}$

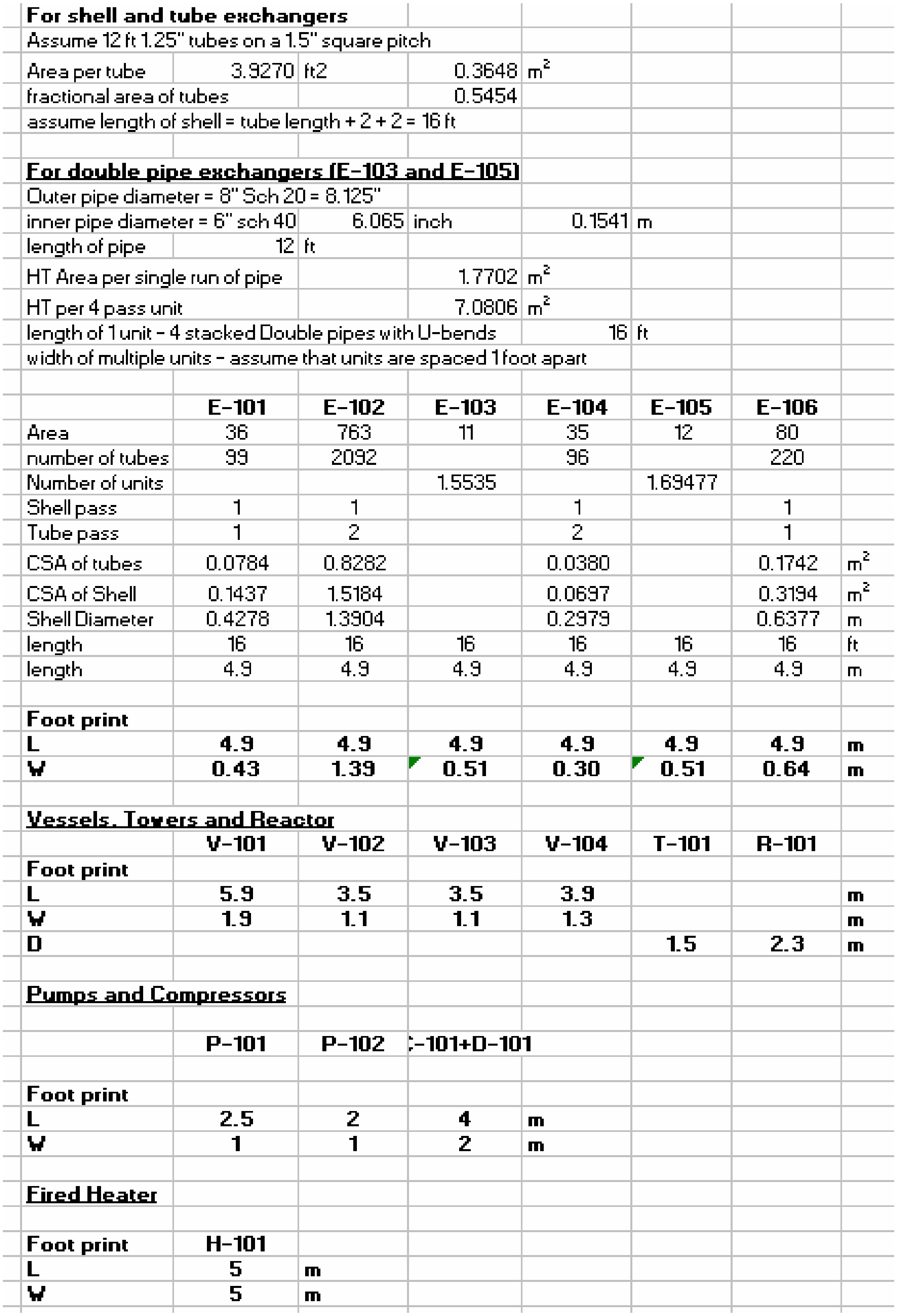
$\boxed{\text{Foot Print } = 1.598 \times 4.877\text{ m}}$

1.18 From Table 1.11 towers and reactors should have a minimum separation of 15 feet or 4.6 meters. No other restrictions apply. See sketch for details.



Solution to Chapter 1 - Problem 18 - many other variations are possible

1.19



For shell and tube exchangers

Assume 12 ft 1.25" tubes on a 1.5" square pitch

Area per tube	3.9270	ft2	0.3648	m^2
fractional area of tubes			0.5454	

assume length of shell = tube length + 2 + 2 = 16 ft

For double pipe exchangers (E-103 and E-105)

Outer pipe diameter = 8" Sch 20 = 8.125"

inner pipe diameter = 6" sch 40	6.065	inch	0.1541	m
length of pipe	12	ft		
HT Area per single run of pipe		1.7702	m^2	
HT per 4 pass unit		7.0806	m^2	
length of 1 unit - 4 stacked Double pipes with U-bends			16	ft

width of multiple units - assume that units are spaced 1 foot apart

	E-101	E-102	E-103	E-104	E-105	E-106	
Area	36	763	11	35	12	80	
number of tubes	99	2092		96		220	
Number of units			1.5535		1.69477		
Shell pass	1	1		1		1	
Tube pass	1	2		2		1	
CSA of tubes	0.0784	0.8282		0.0380		0.1742	m^2
CSA of Shell	0.1437	1.5184		0.0697		0.3194	m^2
Shell Diameter	0.4278	1.3904		0.2979		0.6377	m
length	16	16	16	16	16	16	ft
length	4.9	4.9	4.9	4.9	4.9	4.9	m
Foot print							
L	**4.9**	**4.9**	**4.9**	**4.9**	**4.9**	**4.9**	**m**
W	**0.43**	**1.39**	**0.51**	**0.30**	**0.51**	**0.64**	**m**

Vessels, Towers and Reactor

	V-101	V-102	V-103	V-104	T-101	R-101	
Foot print							
L	**5.9**	**3.5**	**3.5**	**3.9**			**m**
W	**1.9**	**1.1**	**1.1**	**1.3**			**m**
D					**1.5**	**2.3**	**m**

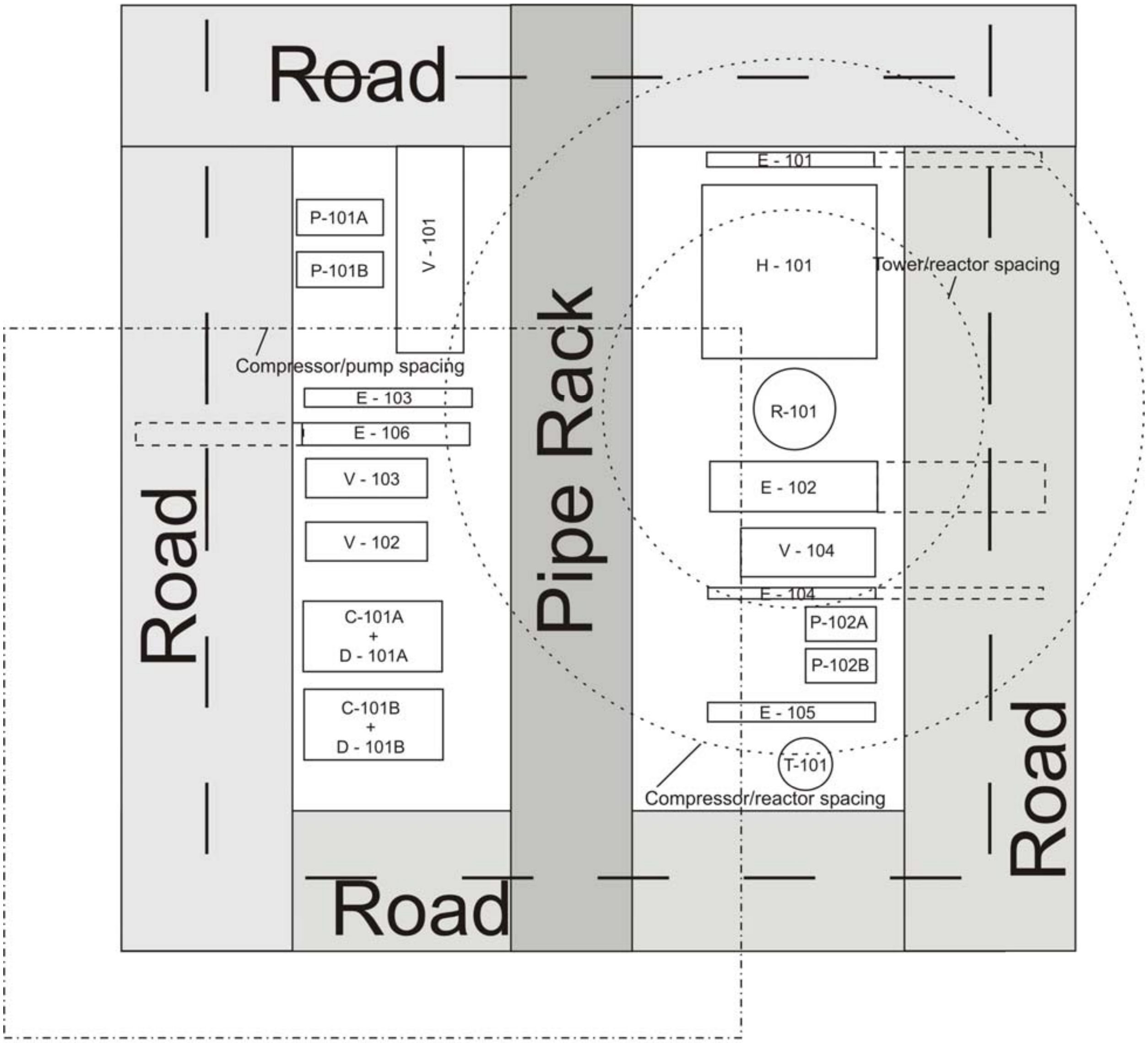
Pumps and Compressors

	P-101	P-102	:-101+D-101	
Foot print				
L	**2.5**	**2**	**4**	**m**
W	**1**	**1**	**2**	**m**

Fired Heater

Foot print	H-101	
L	**5**	**m**
W	**5**	**m**

1.20


Road
Pipe Rack
Road
Road
Road
E - 101
H - 101
Tower/reactor spacing
P-101A
P-101B
V - 101
Compressor/pump spacing
E - 103
E - 106
V - 103
V - 102
C-101A
+
D - 101A
C-101B
+
D - 101B
R-101
E - 102
V - 104
E - 104
P-102A
P-102B
E - 105
T-101
Compressor/reactor spacing

1.21 (a) A temperature (sensing) element (TE) in the plant is connected via a capillary line to a temperature transmitter (TT) also located in the plant. The TT sends an electrical signal to a temperature indicator controller (TIC) located on the front of a panel in the control room.

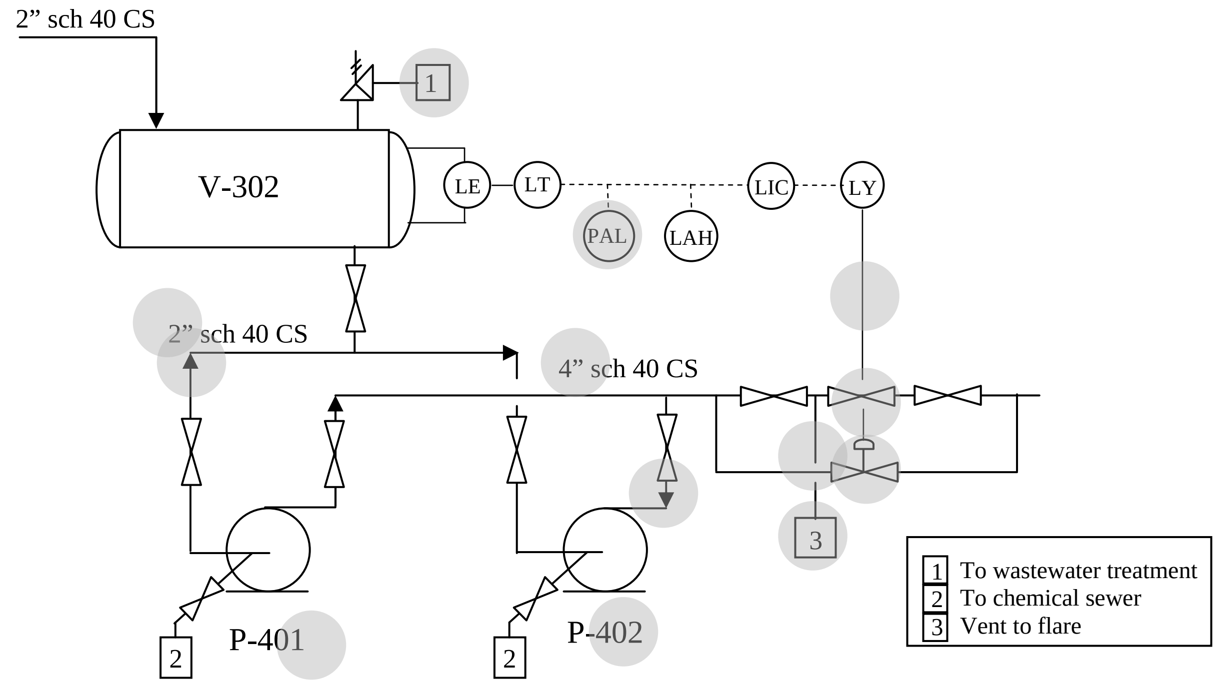
(b) A pressure switch (PS) located in the plant sends an electrical signal to …

(c) A pressure control valve (PCV) located in the plant is connected by a pneumatic (air) line to the valve stem.

(d) A low pressure alarm (PAL) located on the front of a panel in the control room receives an electrical signal from …

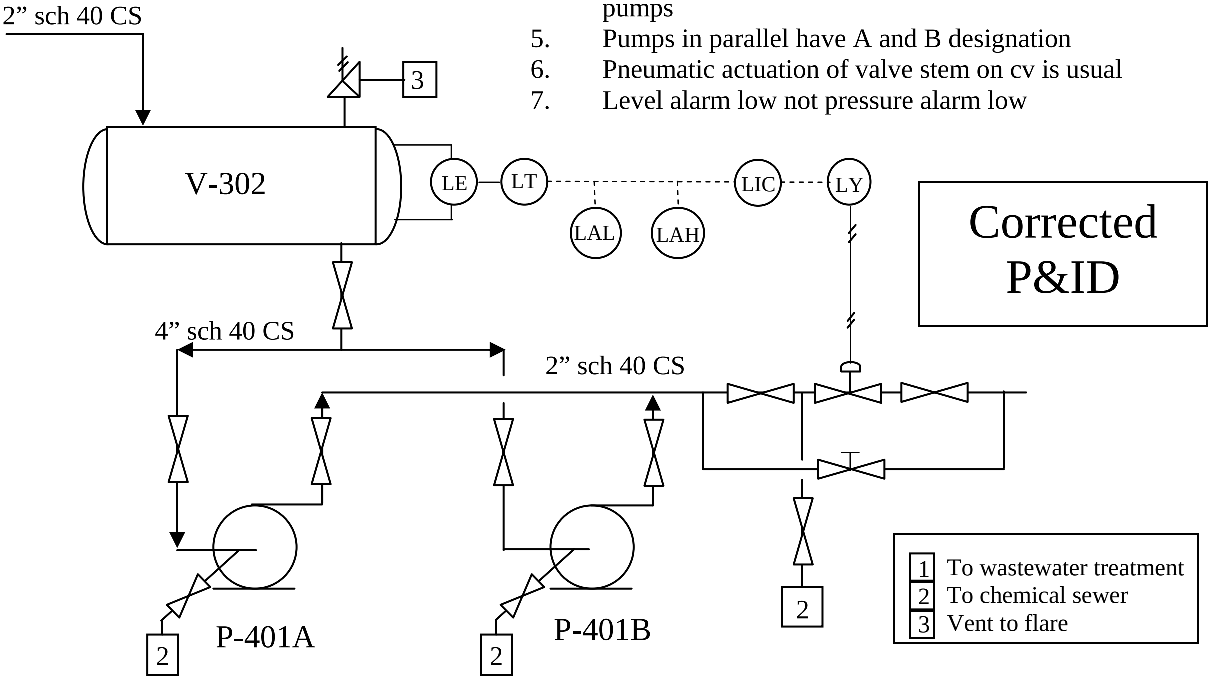
(e) A high level alarm (LAH) located on the front of a panel in the control room receives a signal via a capillary line.

1.22



● = Error

List of Errors
1. Pipe inlet always larger than pipe outlet due to NPSH issues
2. Drains to chemical sewer and vent to flare
3. Double-block and bleed needed on control valve
4. Arrows must be consistent with flow of liquid through pumps
5. Pumps in parallel have A and B designation
6. Pneumatic actuation of valve stem on cv is usual
7. Level alarm low not pressure alarm low



Chapter 2

2.1 The five elements of the Hierarchy of Process Design are:

a. Batch or continuous process
b. Input – output structure of process
c. Recycle structure of process
d. General separation structure of process
e. Heat-exchanger network/process energy recovery

2.2 a. Separate/purify unreacted feed and recycle – use when separation is feasible.
b. Recycle without separation but with purge – when separation of unused reactants is infeasible/uneconomic. Purge is needed to stop build up of product or inerts.
c. Recycle without separation or purge – product/byproduct must react further through equilibrium reaction.

2.3 Batch preferred over continuous when: small quantities required, batch-to-batch accountabilities required, seasonal demand for product or feed stock availability, need to produce multiple products using the same equipment, very slow reactions, and high equipment fouling.

2.4 One example is the addition of steam to a catalytic reaction using hydrocarbon feeds. Examples are given in Appendix B (styrene, acrylic acid.) In the styrene process, superheated steam is added to provide energy for the desired endothermic reaction and to force the equilibrium towards styrene product. In the acrylic acid example, steam is added to the feed of propylene and air to act as thermal ballast (absorb the heat of reaction and regulate the temperature), and it also serves as an anti-coking agent – preventing coking reactions that deactivate the catalyst.

2.5 Reasons for purifying a feed material prior to feeding it to a process include:

a. If impurity foul or poison a catalyst used in the process.
e.g. Remove trace sulfur compounds in natural gas prior to sending to the steam reforming reactor to produce hydrogen.

$$CH_4 + H_20 \rightarrow CO + 3H_2$$

Remove trace sulfur

Platinum catalyst v. susceptible to sulfur poisoning

b. If impurities react to form difficult-to-separate or hazardous products/byproducts.
e.g. Production of isocyanates using phosgene. Production of phosgene is

$$CO + Cl_2 \rightarrow COCl_2$$

The carbon monoxide is formed via steam reforming of CH_4 to give CO + H_2. H_2 must be removed from CO prior to reaction with Cl_2 to form HCl, which is highly corrosive and causes many problems in the downstream processes.

c. If the impurity is present in large quantities then it may be better to remove the impurity rather than having to size all the down stream equipment to handle the large flow of inert material.
e.g. One example is suing oxygen rather than air to fire a combustion or gasification processes. Removing nitrogen reduces equipment size and makes the removal of CO_2 and H_2S much easier because these species are more concentrated.

2.6 IGCC

$$H_2O + C_aH_bS_cO_dN_e + O_2 \rightarrow pCO_2 + qH_2 + rH_2O + sCO + tNH_3 + uH_2S$$

Coal

In modern IGCC plants, coal is partially oxidized (gasified) to produce synthesis gas CO + H_2 and other compounds. Prior to combusting the synthesis gas in a turbine, it must be "cleaned" or H_2S and CO_2 (if carbon capture is to be employed.) Both H_2S and CO_2 are acid gases that are removed by one of a variety of physical or chemical absorption schemes. By removing nitrogen from the air, the raw synthesis gas stream is much smaller making the acid gas removal much easier. In fact, when CO_2 removal is required IGCC is the preferred technology, i.e. the cheapest.

2.7 Ethylebenzene Process

a. Single pass conversion of benzene

Benzene in reactor feed (stream 3) = 226.51 $\frac{\text{kmol}}{\text{h}}$

Benzene in reactor effluent (stream 14) = 177.85 $\frac{\text{kmol}}{\text{h}}$

$$X_{sp} = 1 - \frac{177.85\frac{kmol}{h}}{226.51\frac{kmol}{h}} = 21.5\%$$

b. Single pass conversion of ethylene

Ethylene in reactor feed (stream 2) = 93.0 $\frac{\text{kmol}}{\text{h}}$

Ethylene in reactor effluent (stream 14) = 0.54 $\frac{\text{kmol}}{\text{h}}$

$$X_{sp} = 1 - \frac{0.54\frac{kmol}{h}}{93.0\frac{kmol}{h}} = 99.4\%$$

c. Overall conversion of benzene

Benzene entering process (stream 1) = 97.0 $\frac{\text{kmol}}{\text{h}}$

Benzene leaving process (stream 15 and 19) = 8.38 + 0.17 $\frac{\text{kmol}}{\text{h}}$

$$X_{ov} = 1 - \frac{8.55\frac{kmol}{h}}{97.0\frac{kmol}{h}} = 91.2\%$$

d. Overall conversion of ethylene

Ethylene entering process (stream 2) = 93.0 $\frac{\text{kmol}}{\text{h}}$

Ethylene leaving process (stream 15 and 19) = 0.54 + 0 $\frac{\text{kmol}}{\text{h}}$

$$X_{ov} = 1 - \frac{0.54\frac{kmol}{h}}{93.0\frac{kmol}{h}} = 99.4\%$$

2.8 Separation of G from reactor effluent may or may not be difficult. (a) If G reacts to form a heavier (higher molecular weight) compound then separation may be relatively easy using a flash absorber or distillation and recycle can be achieved easily. (b) If process is to be viable then G must be separable from the product. If inerts enter with G or gaseous by-products are formed then separation of G may not be possible but recycling with a purge should be tried. In either case the statement is not true.

2.9 Pharmaceutical products are manufactured using batch process because:

a. they are usually required in small quantities
b. batch-to-batch accountability and tracking are required by the Food & Drug Administration (FDA)
c. usually standardized equipment is used for many pharmaceutical products and campaigns are run to produce each product – this lends itself to batch operation.

2.10 a. Single pass conversion of ethylbenzene

Ethylbenzene in reactor feed (stream 9) = 512.7 $\frac{\text{kmol}}{\text{h}}$

Ethylbenzene in reactor effluent (stream 12) = 336.36 $\frac{\text{kmol}}{\text{h}}$

$$\text{Single pass conversion} = 1 - \frac{336.36\frac{kmol}{h}}{512.7\frac{kmol}{h}} = 34.4\%$$

b. Overall conversion of ethylbenzene

Ethylbenzene entering process (stream 1) = 180 $\frac{\text{kmol}}{\text{h}}$

Ethylbenzene leaving process (stream 19, 26, 27 & 28) = 3.36 + 0.34 = 3.70 $\frac{\text{kmol}}{\text{h}}$

$$\text{Overall conversion} = 1 - \frac{3.70\frac{kmol}{h}}{180\frac{kmol}{h}} = 97.9\%$$

c. Yield of styrene

Moles of ethylbenzene required to produce styrene = 119.3 $\frac{\text{kmol}}{\text{h}}$

Moles of ethylbenzene fed to process (stream 1) = 180 $\frac{\text{kmol}}{\text{h}}$

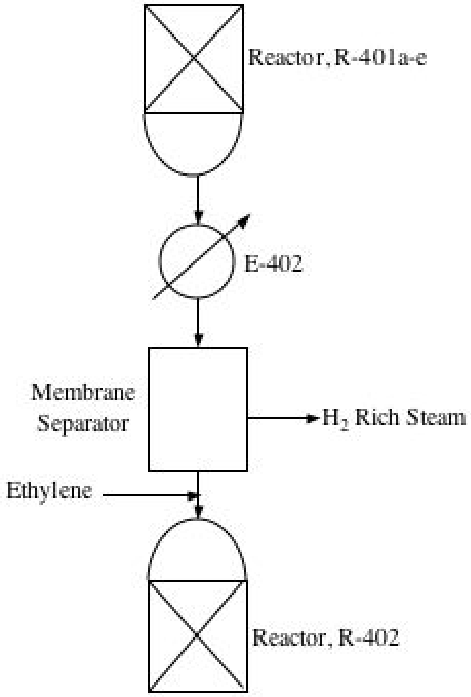
$$\text{Yield} = \frac{119.3\frac{kmol}{h}}{180\frac{kmol}{h}} = 66.3\%$$

Possible strategies to increase the yield of styrene are

(i) Increase steam content of reactor feed – this pushes the desired equilibrium reaction to the right.

(ii) Increasing the temperature also pushes the equilibrium to right but increases benzene and toluene production.

(iii) Remove hydrogen in effluent from each reactor – this will push the equilibrium of the desired reaction to the right and reduce the production of toluene from the third reaction – use a membrane separator, shown on following page.


Reactor, R-401a-e
E-402
Membrane Separator
H_2 Rich Steam
Ethylene
Reactor, R-402