Process

First semester 2025-2026

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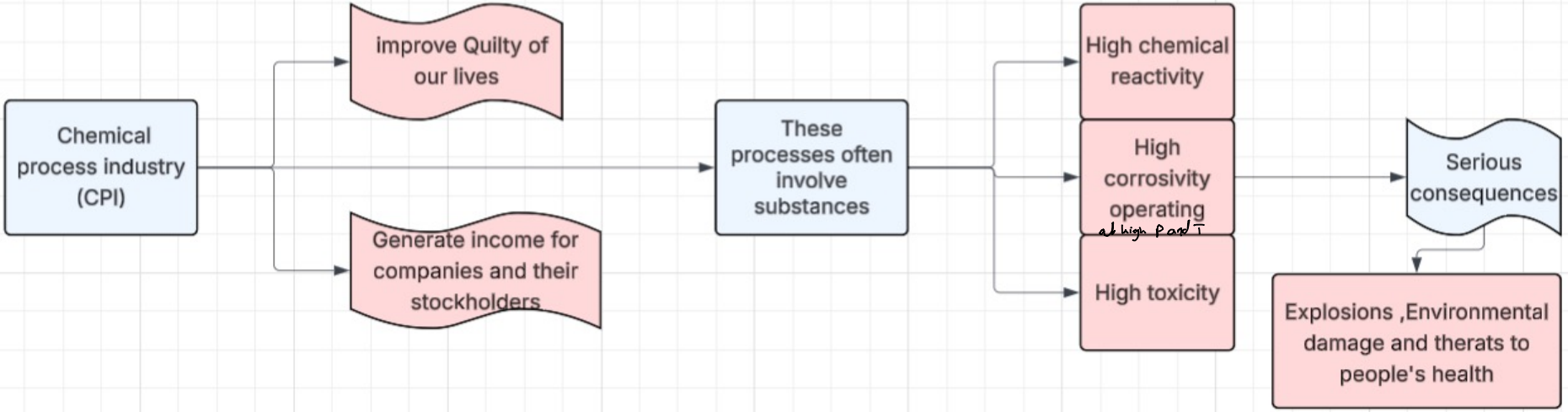
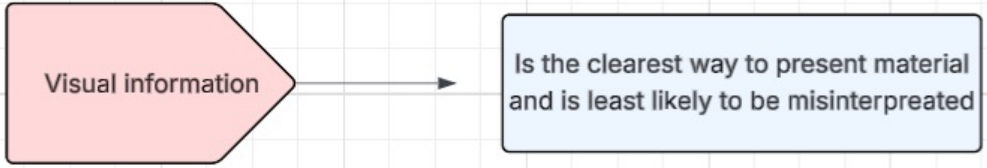
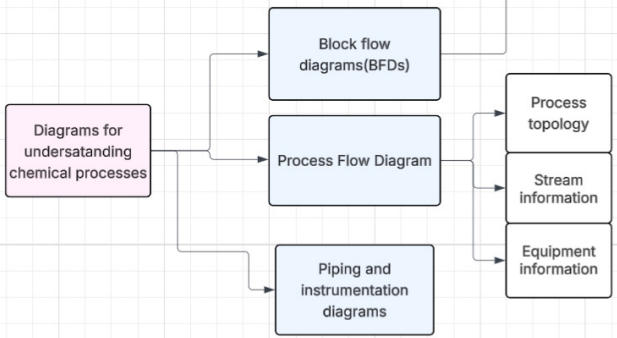
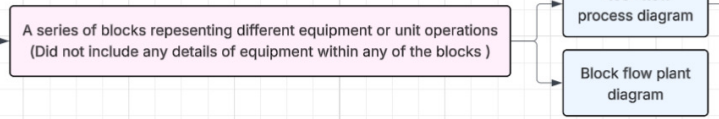
Madar Team

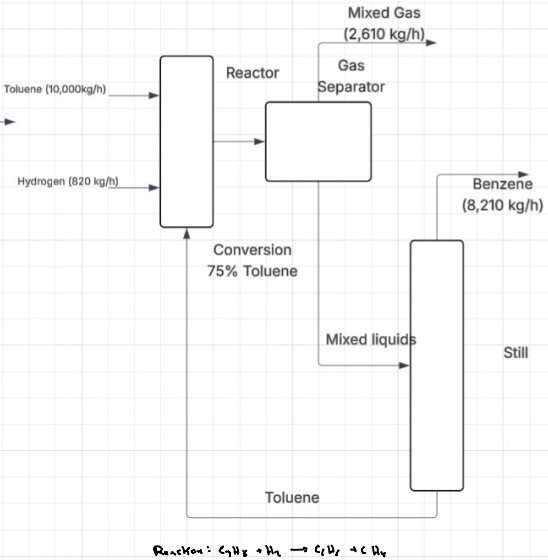

Chemical process industry (CPI)
improve Quilty of our lives
Generate income for companies and their stockholders
These processes often involve substances
High chemical reactivity
High corrosivity operating at high Pand T
High toxicity
Serious consequences
Explosions ,Environmental damage and therats to people's health


Visual information
Is the clearest way to present material and is least likely to be misinterpreated


Diagrams for undersatanding chemical processes
Block flow diagrams(BFDs)
Process Flow Diagram
Piping and instrumentation diagrams
Process topology
Stream information
Equipment information


A series of blocks repesenting different equipment or unit operations
(Did not include any details of equipment within any of the blocks)

Block flow process diagram
Block flow plant diagram


Toluene (10,000kg/h)
Hydrogen (820 kg/h)
Reactor
Mixed Gas (2,610 kg/h)
Gas Separator
Benzene (8,210 kg/h)
Conversion 75% Toluene
Mixed liquids
Still
Toluene
Reaction: C7H8 + H2 → C6H6 + CH4

Equipment

Flow summary table

Electricity

Compressed air

cooling water

refrigerated water

steam

condensate return

inert gas for blankieting

chemical sewer

waste water treatment

flares

Utility

Control loops

Process flow diagram (PFD)

Is a complex diagram are drawn on large sheets of paper

Information flags

Dual role

Needed in the plant design

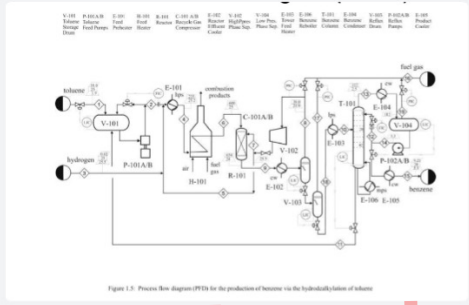
Analysis of operating problems

Basic information provided by a PFD

Process topology

Stream informatin

Equipment information



highly detailed and precise

Piping and Instrumentation Diagram (P&ID)

piping, process, instrumentation, and other diagrams

Guide for those who will be responsible

Project engineers will develop plant and construction schedules

Mechanical engineers and Civil engineers will design and install pieces of equipment

Instrument engineers will specify, instal, and check control systems

Piping engineers will develop plant layout and elevation drawings

To constract the plant and specify all mechanical details

Mechanical Flow Diagram (MFD)

P&ID

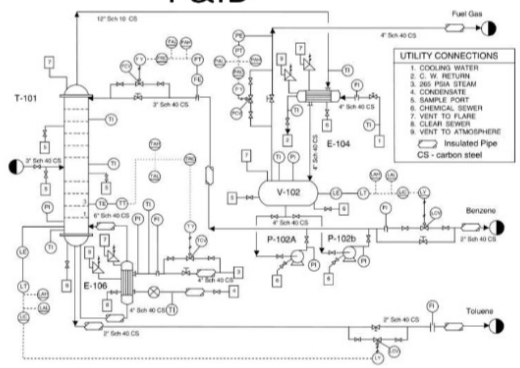


Figure 1.7 : Piping and Instrumentation Diagram for Benzene Distillation (adapted from Kauffman, D. Flow Sheets and Diagrams," AIChE Modular Instruction, Series G: Design of Equipment, series editor J. Beckman, AIChE, New York, 1986, vol 1, Chapter G.1.5, AIChE copyright © 1986 AIChE, all rights reserved)

Chapter 1 Diagrams for understanding chemical processes

Flow diagrams

Laboratory stage ⟶ Design stage ⟶ construction stage ⟶ Plant operation

تجارب المختبر — رسم مخططات — توجيه الكتابة — ادارة

1.1 Block Flow Diagrams (BFDs)

First step is convert word problem to block diagram

Consist ① series of blocks representing equipment or unit operations

② Arrows masses ⟶ streams input and output

يحتوي على معلومات مثل

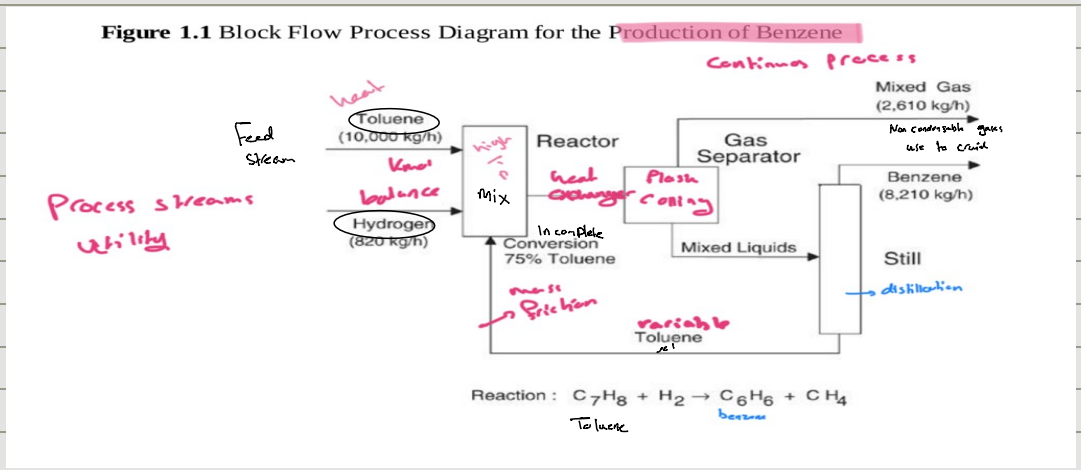
T, P, Conversion, y, Flow rates, Chemical comp

BFD ↗ block flow process diagram

↘ block flow plant diagram

Block flow process diagram

Figure 1.1 Block Flow Process Diagram for the Production of Benzene



reaction ⟶ separation ⟶ distillation

① by blocks

② line ⟶ direction of flow

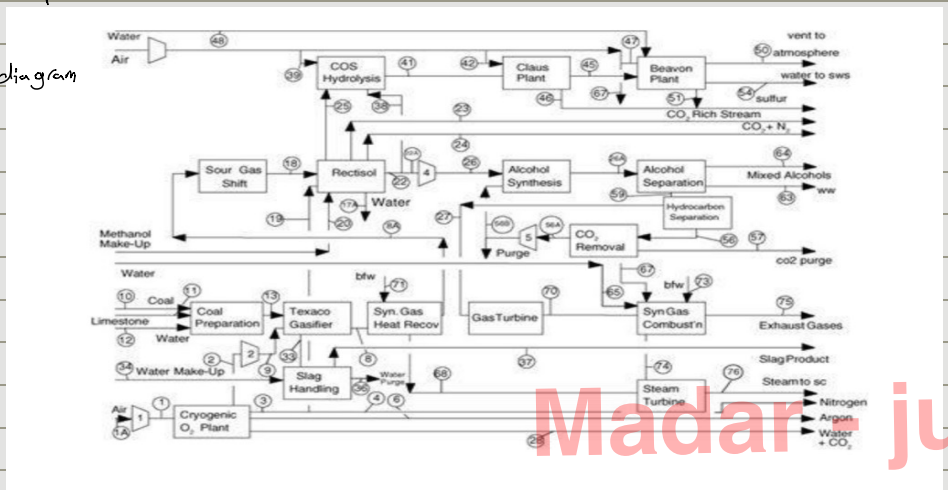
③ flow left ⟶ right

④ light (gases) top heavy stream (liquid, solids) bottom

⑤ 1/1

⑥ simplified material balance provided.

Block flow plant diagram



	Block flow process diagrams	Block flow plant Diagram
orientation	Shows the process within a specific area	provides an overall view of the plant, products and key operating areas
Process design	used to evaluate alternatives within a specific process	used to understand interactions between different processes
Simplification	Basic information for each unit operation without equipment details	General and clear information without detailed specification

2. PFD containing following information

(1) Major equipment each Equipment have a number and descriptive name

(2) process flow stream will be shown and identified by a number. This data will be either displayed directly on the PFD or included in summary table

(3) Utility streams

(4) Basic control loop

Process Topology (basic structure)

Shows the overall arrangement of major equipment and how process streams connect between them

main equipment (reactor, heat exchangers, separator, distillation column)

flow connections (feed, recycle, and product stream)

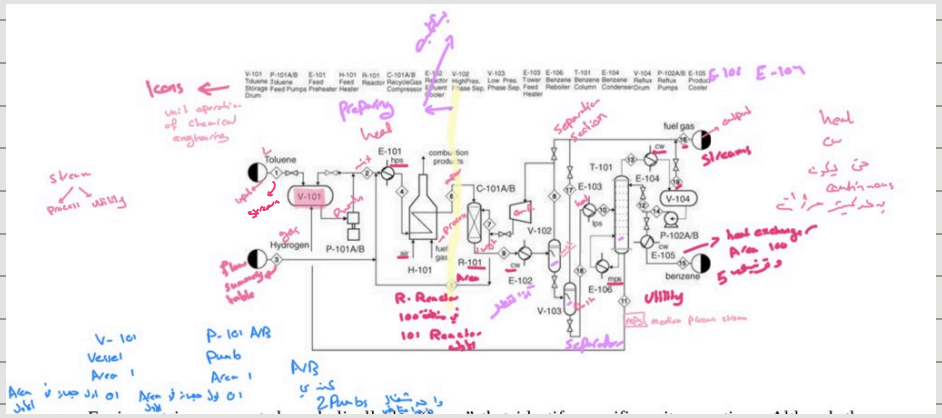


Table 1.2 Conventions Used for Identifying Process Equipment

Process Equipment	General Format XX-YZZ A/B → Spare
	XX are the identification letters for the equipment classification
	C - Compressor or Turbine
	E - Heat Exchanger
	H - Fired Heater
	P - Pump
	R - Reactor
	T - Tower
	TK - Storage Tank
	V - Vessel
	Y designates an area within the plant
	ZZ is the number designation for each item in an equipment class
	A/B identifies parallel units or backup units not shown on a PFD
Supplemental Information	Additional description of equipment given on top of PFD

Ex P. 101 A/B

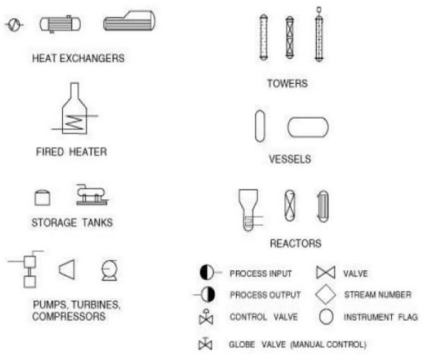
P: Pump

1 → 100 Area 100

01 → specific pump is number 01 in unit 100

A/B → Two identical Pumps P-101 A and P-101 B. One pump will be operating while the other is idle

Figure 1.4 Symbols for Drawing Process Flow Diagrams



Process Streams
All conventions shown in Table 1.1 apply.
Diamond symbol located in flow lines.
Numerical identification (unique for that stream) inserted in diamond.
Flow direction shown by arrows on flow lines.

Utility Streams	
lps	Low-Pressure Steam: 3–5 barg (sat) ‡
mps	Medium-Pressure Steam: 10–15 barg (sat) ‡
hps	High-Pressure Steam: 40–50 barg (sat) ‡
htm	Heat Transfer Media (Organic): to 400°C
cw	Cooling Water: From Cooling Tower 30°C Returned at Less Than 45°C†
wr	River Water: From River 25°C Returned at Less Than 35°C
rw	Refrigerated Water: In at 5°C Returned at Less Than 15°C
rb	Refrigerated Brine: In at –45°C Returned at Less Than 0°C
cs	Chemical Waste Water with High COD
ss	Sanitary Waste Water with High BOD, etc.
el	Electric Heat (Specify 220, 440, 660V Service)
ng	Natural Gas
fg	Fuel Gas
fo	Fuel Oil
fw	Fire Water

‡These pressures are set during the preliminary design stages and typical values vary within the ranges shown.
†Above 45°C, significant scaling occurs.

Required Information Essential information
Stream Number
Temperature (°C)
Pressure (bar)
Vapor Fraction
Total Mass Flowrate (kg/h)
Total Mole Flowrate (kmol/h)
Individual Component Flowrates (kmol/h)
Optional Information اطلاعات تکمیلی
Component Mole Fractions
Component Mass Fractions
Individual Component Flowrates (kg/h)
Volumetric Flowrates (m^3/h)
Significant Physical Properties Density Viscosity Other
Thermodynamic Data Heat Capacity Stream Enthalpy K-values
Stream Name

A Portion of Table 1.5

Stream Number	1	2	3	4	5	6	7	8	9	10
Temperature (°C)	25	59	25	225	41	600	41	38	654	90
Pressure (bar)	1.90	25.8	25.5	25.2	25.5	25.0	25.5	23.9	24.0	2.6
Vapor Fraction	0.0	0.0	1.00	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Mass Flow (tonne/h)	10.0	13.3	0.82	20.5	6.41	20.5	0.36	9.2	20.9	11.6
Mole Flow (kmol/h)	108.7	144.2	301.0	1204.4	758.8	1204.4	42.6	1100.8	1247.0	142.2
Component Mole Flow (kmol/h)										
Hydrogen	0.0	0.0	286.0	735.4	449.4	735.4	25.2	651.9	652.6	0.02
Methane	0.0	0.0	15.0	317.3	302.2	317.3	16.95	438.3	442.3	0.88
Benzene	0.0	1.0	0.0	7.6	6.6	7.6	0.37	9.55	116.0	106.3
Toluene	108.7	143.2	0.0	144.0	0.7	144.0	0.04	1.05	36.0	35.0

Example 1.2

Overall material balance for the benzene process

input stream 1 (toluene feed) 10

stream 3 (hydrogen feed) 0.82

10.82×10^3 kg/h

output stream 15 (product benzene) 8.21

stream 16 (fuel gas) 2.61

10.82×10^3 kg/h

input = output

Conversion is defined as $\epsilon = \frac{\text{Benzene produced}}{\text{total toluene introduced}}$

Example 1.3

R-101

input stream 2 & 7

output stream 9

total toluene introduced = 144 + 0.04 = 144.04 Kmol/h

Benzene produced = 116 − 7.6 − 0.37 = 108.03 Kmol/h

$\epsilon = \frac{108.03}{144.04} = 0.75$

Piping and Instrument Diagram (P and ID).

Location of Instrumentation	
◯	Instrument Located in Plant
⊖	Instrument Located on Front of Panel in Control Room
⊖ (dashed)	Instrument Located on Back of Panel in Control Room

Meanings of Identification Letters (XYY)		
	First Letter (X)	*Second or Third Letter (Y)*
A	Analysis	Alarm
B	Burner Flame	
C	Conductivity	Control
D	Density or Specific Gravity	
E	Voltage	Element
F	Flowrate	
H	Hand (Manually Initiated)	High
I	Current	Indicate
J	Power	
K	Time or Time Schedule	Control Station
L	Level	Light or Low
M	Moisture or Humidity	Middle or Intermediate
O		Orifice
P	Pressure or Vacuum	Point
Q	Quantity or Event	
R	Radioactivity or Ratio	Record or print
S	Speed or Frequency	Switch
T	Temperature	Transmit
V	Viscosity	Valve, Damper, or Louver
W	Weight	Well
Y		Relay or Compute
Z	Position	Drive

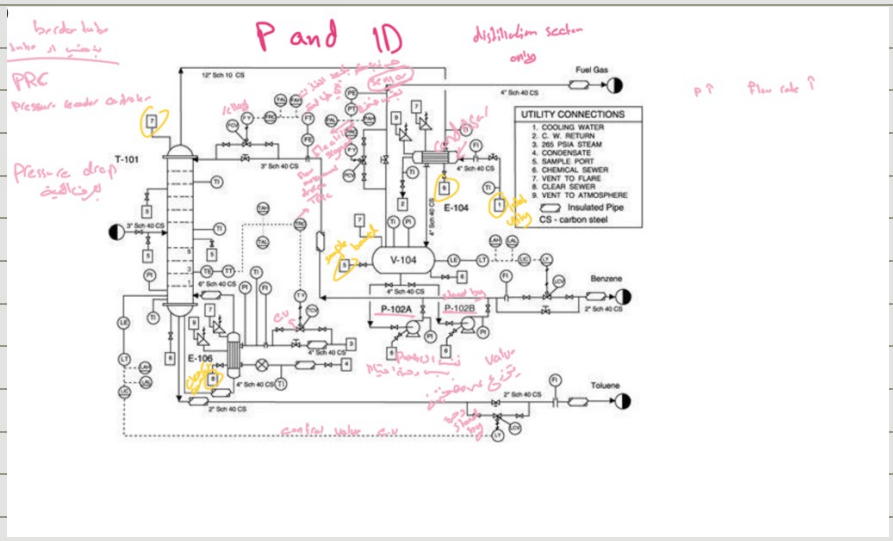
Identification of Instrument Connections	
———	Capillary
—//	Pneumatic
.....	Electrical

Piping and Instrumentation Diagram (P & ID)

Mechanical Flow digaram (MFD)

Provides information needed be engineers to begin planning for the construction of the plant.

	PFD	P & ID
Purpose	To understand the overall chemical process.	To construct the plant and specify all mechanical details.
Content	Material flows, main equipment, general operating conditions	piping, valves, control instruments, instrumentation details
level of detail	General and Simplified	Highly detailed and precise
usage	Conceptual design and economic analysis	Mechanical and implementation design
prepared by	operating company or process engineers.	Mechanical and control design engineers.



The P & ID is the last stage of process design and serves as a guide for those who will be responsible for the final design and construction

- Mechanical engineers and civil engineers will design and install pieces of equipment.
- Instrument engineers will specify install, and start control systems.
- Piping engineers will develop plant layout and elevation drawings.
- Project engineers will develop plant and construction schedules.

Operators Training

Start up procedures

Operation and monitoring.

Additional diagrams

→ Utility flowsheet

Vessel sketches and equipment diagrams

logic ladder and wiring diagrams

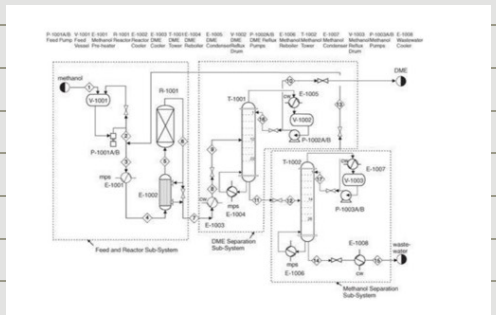
Site plans and structural support drawings.

Dimethyl Ether (DME)

Feed and reactor section

DME purification section

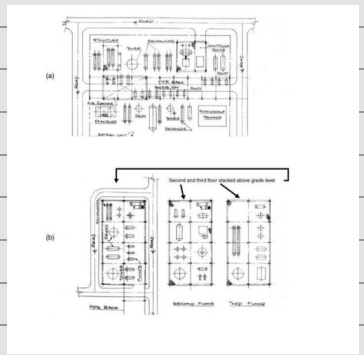
Methanol separation and recycle section



Preliminary plot plans

→ Grade-level horizontal, linear arrangement

→ Structure-Mounted vertical arrangement



What are the three principal types of diagrams used by process engineers to describe the flow of chemicals in a process? On which of these diagrams would you expect to see the following items:

a. the temperature and pressure of a process stream

b. an overview of a multiple-unit process

1.

c. a major control loop

d. a pressure indicator

e. a pressure-relief valve

A problem has occurred in the measuring element of a level-indicating controller in a batch reactor.

2.

To what principal diagram should you refer to in order to troubleshoot the problem?

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

Which of the principal diagrams should be used to do the following:

a. Determine the number of trays is a distillation column?

b. Determine the top and bottom temperatures in a distillation column?

c. Validate the overall material balance for a process?

7.

d. Check the instrumentation for a given piece of equipment in a "pre-start-up" review?

e. Determine the overall material balance for a whole chemical plant?

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

a)The efficiency of a fired heater had been specified incorrectly as 92% instead of 82%.

b. A waste process stream flowrate (sent to a sludge pond) was calculated incorrectly and is now 30% greater than before.

c. It has been decided to add a second (backup) drive for an existing compressor.

d. The locations of several control valves have changed to allow for better operator access.

10. During a retrofit of an existing process, a vessel used to supply the feed pump to a batch reactor has been replaced because of excessive corrosion. The vessel is essentially identical to the original one, except it is now grounded differently to reduce the corrosion. If the function of the vessel (namely to supply liquid to a pump) has not changed, answer the following questions:

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

a. Should the new vessel have a new equipment number, or should the old vessel number be usedagain? Explain your answer.

b.On which diagram or diagrams (BFD, PFD, or P&D) should the change in the grounding setup be noted?

11. Draw a section of a P&ID diagram for a vessel receiving a process liquid through an insulated 4″ sch 40 pipe. The purpose of the vessel is to store approximately 5 minutes of liquid volume and to provide “capacity” for a feed pump connected to the bottom of the pump using a 6″ sch 40 pipe. The diagram should include the following features:

a. The vessel is numbered V-1402 and the pump(s) are P-1407 A/B.

b. The discharge side of the pump is made of 4″ sch 40 carbon steel pipe and all pipe is insulated.

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.

What do the following symbols (as seen on a P&ID) indicate?

21.

(a) TE — TT ····· TIC

(b) PS ·····

(c) PCV

(d) PAL ·····

(e) LAH ——

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.

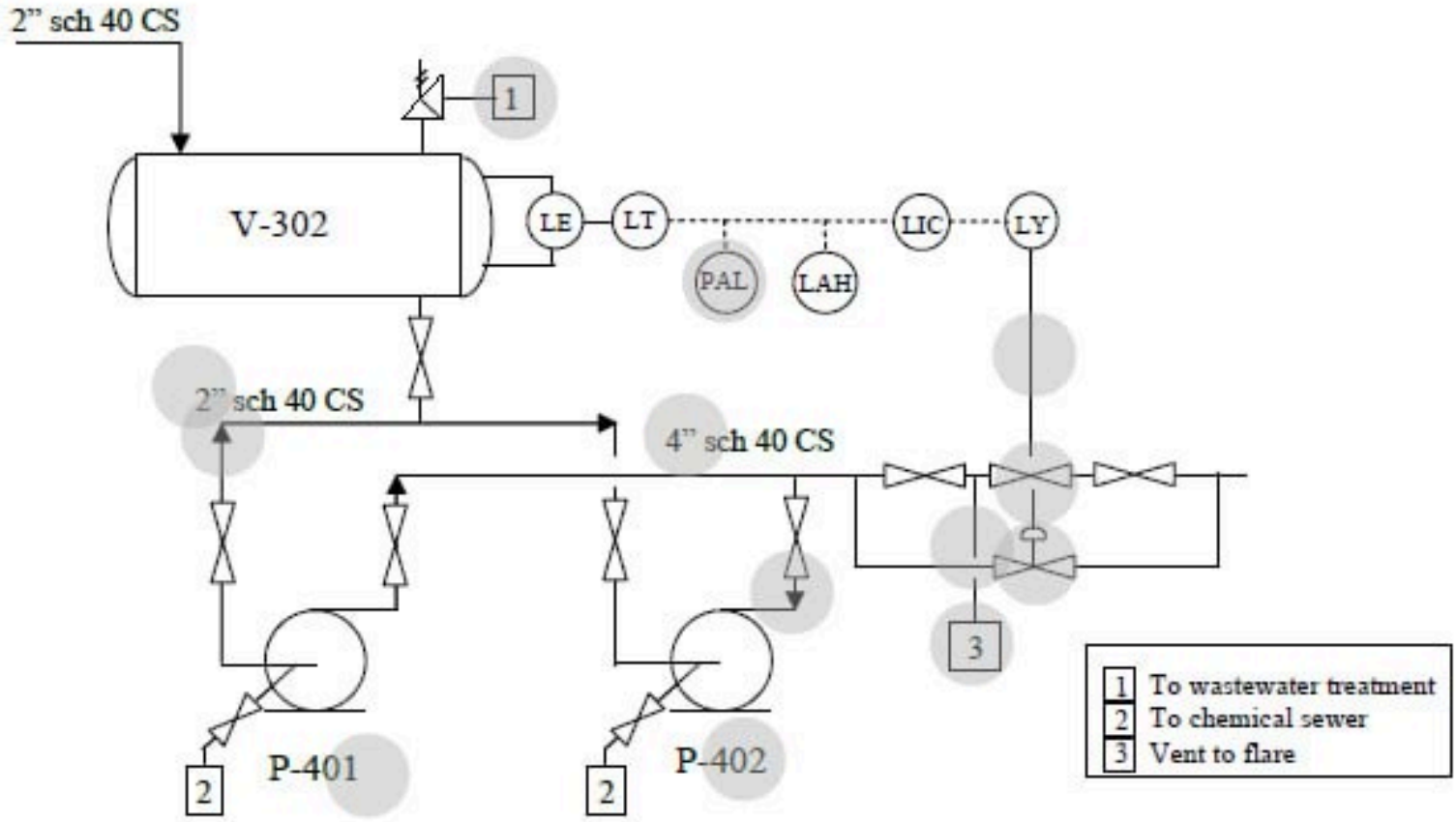
(b) A pressure switch (PS) located in the plant sends an electrical signal to …

(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.

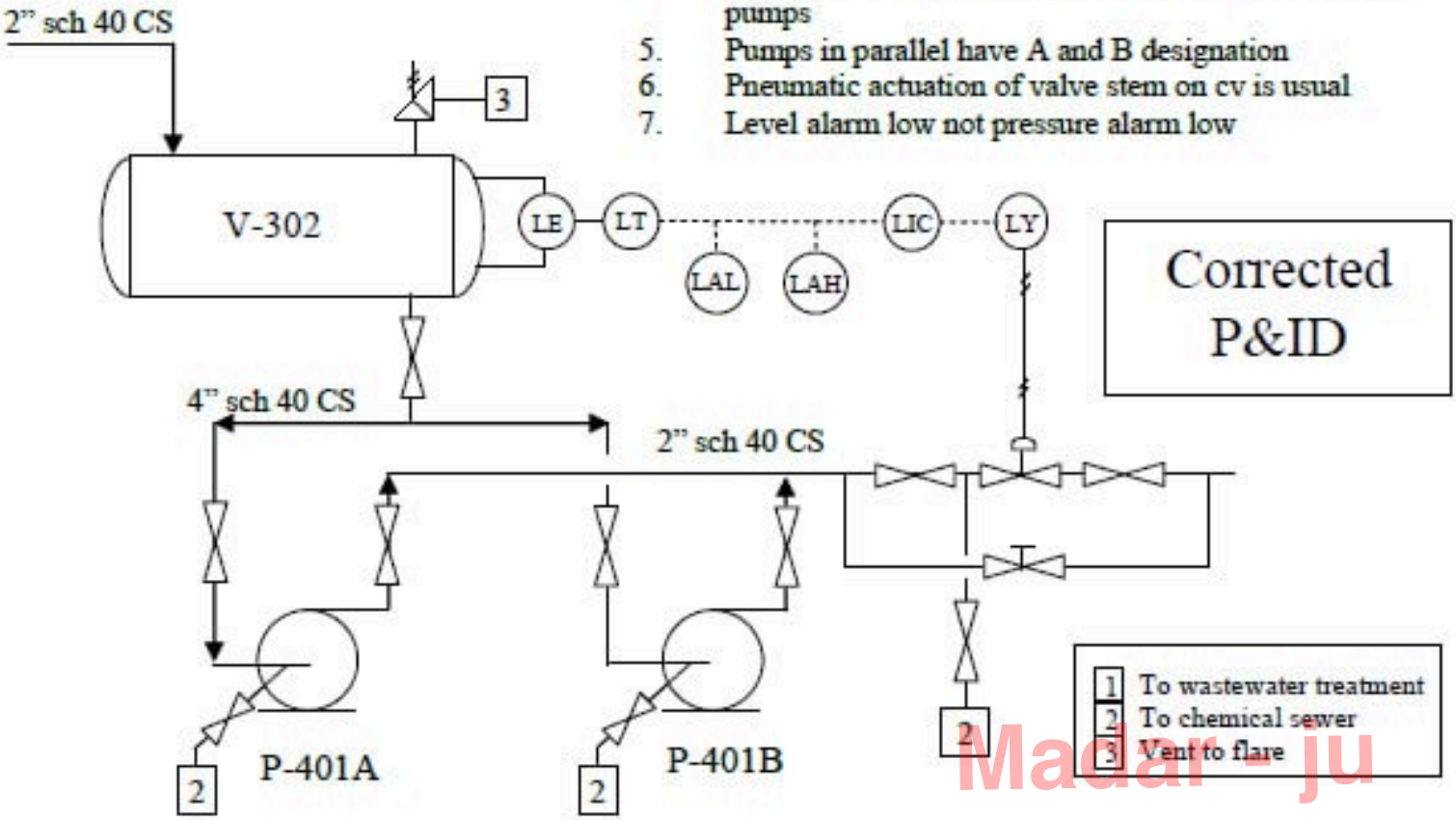
1.22



= Error

List of Errors

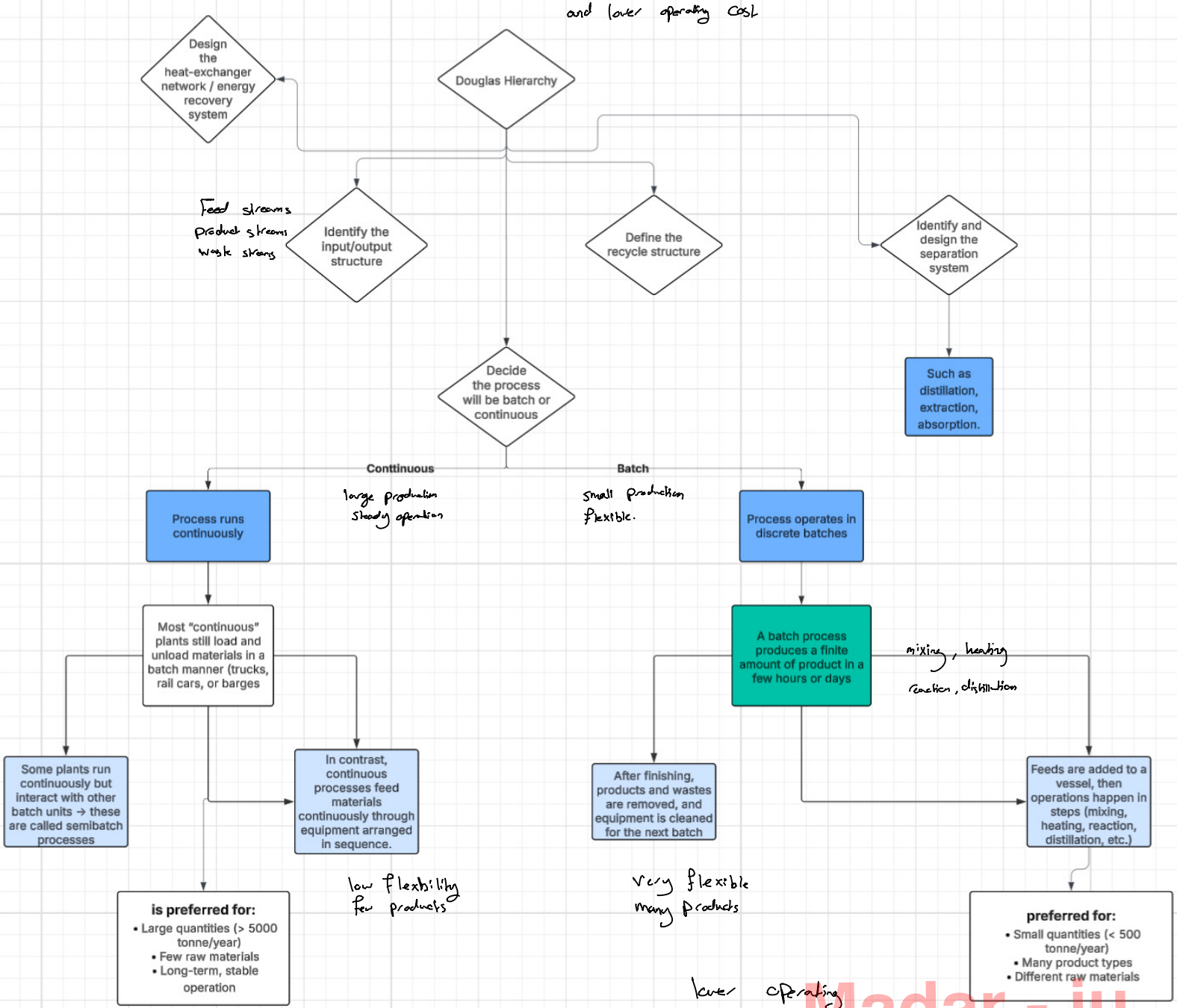
1. Pipe inlet always larger than pipe outlet due to NPSH issues
2. Drains to chemical sewer and vent to flare
3. Double-block and bleed needed on control valve
4. Arrows must be consistent with flow of liquid through pumps
5. Pumps in parallel have A and B designation
6. Pneumatic actuation of valve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



Corrected P&ID

Ch2

* Batch process are preferred for small production rates and high flexibility, while continuous process are favored for large scale production and lower operating cost



But laboratory experiments do not easily reveal problems like

- Trace impuritie
- Material handling difficulties
- Multiple reaction steps interactions

Inputs = feed streams
outputs = product or waste streams

Process Concept Diagram

The first step in evaluating a process route is constructing a process concept diagram

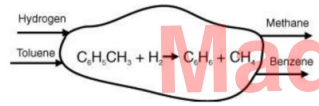
This diagram uses reaction stoichiometry to identify feeds and products

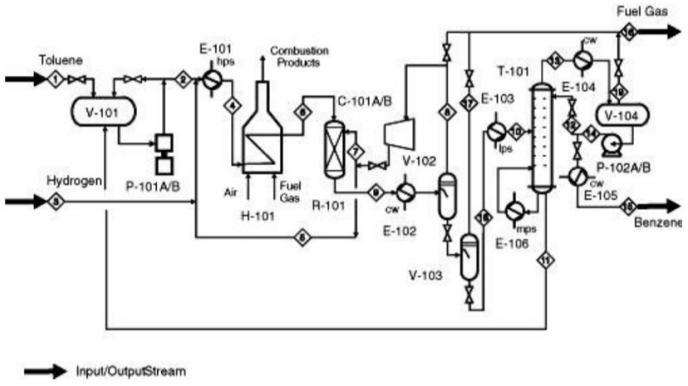
First, identify the chemical reaction(s) occurring in the process

Steps to Build a Process Concept Diagram

- Draw a "cloud" representing the whole process and write the reaction stoichiometry inside it
- Draw reactant streams entering from the left; one stream for each reactant
- Draw product streams leaving from the right; one for each product.
  Include all side reactions and unwanted by-products

Figure 2.1 Input/Output Structure of the Process Concept Diagram for the Toluene Hydrodealkylation Process





The stoichiometry of the reaction, the amount of benzene and methane produced should be equal on a mole basis

Benzene produced = benzene leaving − benzene entering

= 2.85 + 105.2

= 108.05 kmol/h

Methane produced = methane leaving − methane entering

= 15 − 12305

= 108.05 kmol/h

* Factors in Analyzing the input/output Structure

1. Chemicals that enter but are not consumed.

They may be catalysts / solvents / inhibitors / inerts impurities in the feed / materials added to control reaction rate or temperature.

2. Chemicals leaving the process.

have entered in one of feed streams

or have been produced by a chemical reaction.

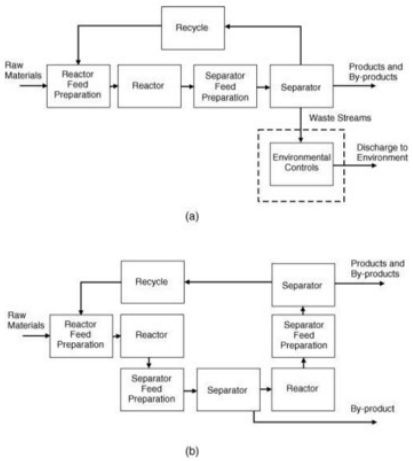
3. Utility streams vs process streams.

Utility streams (steam, cooling water, fuel, electricity) are handled differently they do not mix with process streams.

Their main roles are

adding / removing heat

Providing work or power.



① Reactor Feed preparation block

In most cases, feed chemicals enter the process from storage tanks these feeds usually are not at the right conc., T, P needed for the reactor to operate efficiently

The purpose of this block is to adjust the feed conditions so they meet the reactor's requirements

* Typical operations include

Heating or cooling

Pressurizing or depressurizing

Mixing with other components

Removing impurities.

② Reactor block

The stream leaving the reactor contains

Desired products

unused reactants

Undesired by product (from side reactions)

3. Separator Feed Preparation block

The reactor effluent is usually not in the right temperature or pressure for effective separation

Operations often include

Cooling Compression

Heating

Pressure drop

4. Separator block

This block performs the physical separation of

products Waste streams

by-products Unreacted materials

Common separation methods include

Distillation / Adsorption / liquid-liquid extraction.

5. Recycle block

Unreacted feed materials are collected and sent back to the reactor to be used again. This improves: Efficiency, Economics of the process

Equipment typically includes

Pump or compressor

Heat exchanger (sometimes)

6. Environmental control block

All chemical process produce waste streams (gases, liquids, solids). These must be treated before release to the environment.

Waste may include:

Unreacted materials

Side product chemicals

Impurities in feed

Emissions from energy generation (ex, burning high sulfur fuel produces SO_2)

Heuristics For handling Feed impurities

(1) When impurities are small (5%-20%) and do not react ⟶ Do not separate

(2) If separation of impurities is difficult ⟶ Do not separate

(3) If impurities poison or foul the catalyst ⟶ purify the feed.

(4) If impurities form hazardous or hard-to-separate products ⟶ purify the Feed.

(5) If impurities is present in large amounts ⟶ purify the feed (usually)

1.What are the five elements of the hierarchy of process design?

2.What are the three types of recycle structures possible in a chemical process? Explain when each is used.

3. Give three criteria for choosing a batch process as opposed to a continuous process.

4.When would one purposely add an inert material to a feed stream? Illustrate this strategy with an example,and explain the advantages (and disadvantages)of ding this .

5.In general, when would one puirfy of a material prior to feeding it to a process unti? Give at least one example for each case you state .

Chapter 2

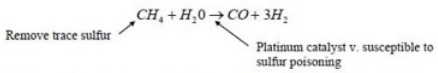
2.1 The five elements of the Hierarchy of Process Design are:
a. Batch or continuous process
b. Input – output structure of process
c. Recycle structure of process
d. General separation structure of process
e. Heat-exchanger network/process energy recovery

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.



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.

7.The production of ethylbenzene is described in Appendix B, project B.2. From the PFD (Figure B.2.1) and accompanying stream table (Table B.2.1), determine the following:

a. The single-pass conversion of benzene

b. The single-pass conversion of ethylene

c. Overall conversion of benzene

d. Overall conversion of ethylene

Suggest two strategies to increase the overall conversion of ethylene and discuss their merits.

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 $\frac{\text{kmol}}{\text{h}}$

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

9. Most pharmaceutical products are manufactured using batch processes. Give at least three reasons why this is so.

The formation of styrene via the dehydrogenation of ethylbenzene is a highly endothermic reaction. In addition, ethylbenzene may decompose to benzene and toluene and also may react with hydrogen to form toluene and methane:

(B.2.1)

$$\underset{\text{ethylbenzene}}{C_6H_5C_2H_5} \xrightarrow{1} \underset{\text{styrene}}{C_6H_5C_2H_3} + \underset{\text{hydrogen}}{H_2}$$

(B.2.2)

$$\underset{\text{ethylbenzene}}{C_6H_5C_2H_5} \xrightarrow{3} \underset{\text{benzene}}{C_6H_6} + \underset{\text{ethylene}}{C_2H_4}$$

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.

10. (B.2.3)

$$\underset{\text{ethylbenzene}}{C_6H_5C_2H_5} + \underset{\text{hydrogen}}{H_2} \xrightarrow{4} \underset{\text{toluene}}{C_6H_6CH_3} + \underset{\text{methane}}{CH_4}$$

This process is presented in Appendix B as project B.3. From the information given in Appendix B, determine the following:

a. the single-pass conversion of ethylbenzene.

b. the overall conversion of ethylbenzene.

---

c. the yield of styrene.

Suggest one strategy to increase the yield of styrene, and sketch any changes to the PFD that this strategy would require.

There are two technically viable routes to the production of a hydrocarbon solvent, *S*, starting with feed material *A*. Route 1 uses a disproportionation reaction, in which feed material *A* is converted to the desired solvent *S* and another solvent *R*, both of which are marketable products. Route 2 starts with the same chemical *A*, but uses a hydrodealkylation reaction to produce the desired solvent. The reaction schemes for each process are shown below.

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 $\frac{\text{kmol}}{\text{h}}$

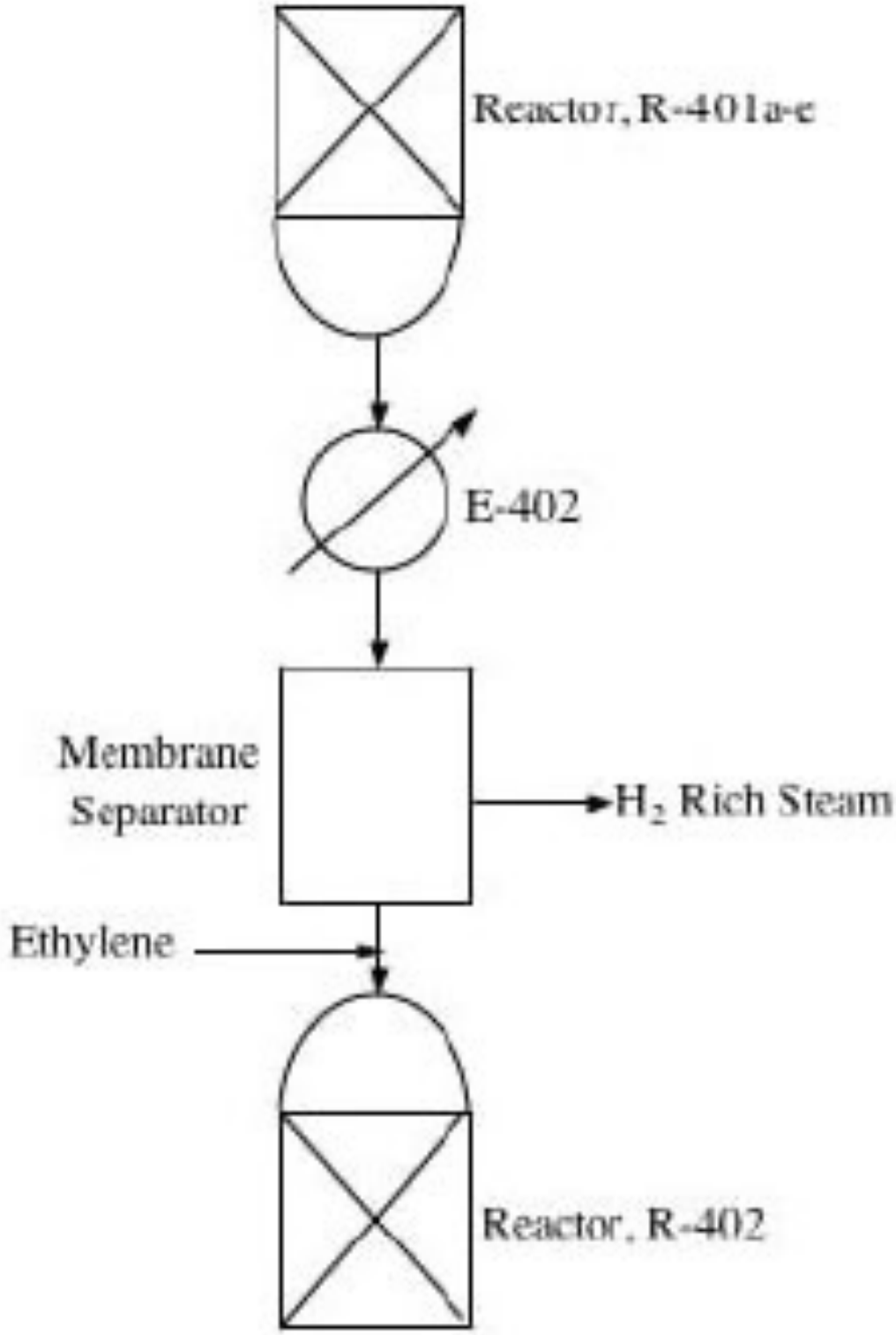
$$\text{Yield} = \frac{119.3\frac{kmol}{h}}{180\frac{kmol}{h}} = 66.3\%$$

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

11.

Route 1 $2A \rightarrow S + R$
Route 2 $A + H_2 \rightarrow S + CH_4$

Assuming that pure *A* is fed to the process, the solvents *S* and *R* are separable by simple distillation, and both are much less volatile than either methane or hydrogen, sketch PFDs for Routes 1 and 2. Which process do you think will be more profitable? Explain your reasoning and assumptions.

When considering the evolution of a process flowsheet, it was noted that there are three forms of recycle structure for unused reactants, given as a–c below. For each case, carefully explain under what conditions you would consider or implement each strategy.

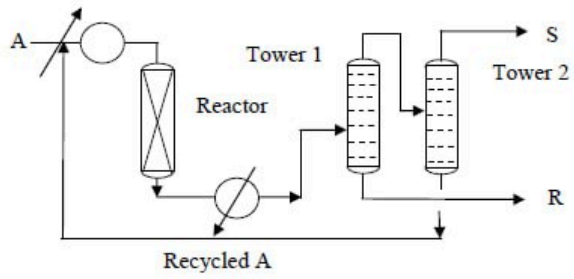
2.11 Route 1: $2A \rightarrow S + R$

Key features are that no light components (non-condensables) are formed and only one reactant is used. Therefore, separation of A, R, and S can take place using distillation columns.

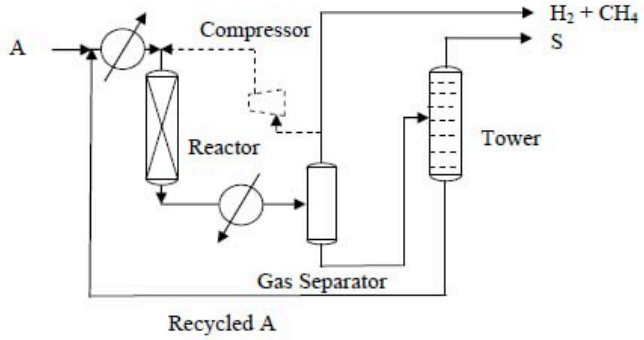
Route 2: $A + H_2 \rightarrow S + CH_4$

Unlike Route 1, this process route requires separation of the non-condensables from A and S. If hydrogen is used in great excess (as with the toluene HDA process), then a recycle and purge of the light gas stream will be required. Otherwise, if hydrogen conversion is high, the unreacted hydrogen along with the methane may be vented directly to fuel gas.

Route 1 – PFD sketch



Route 2 – PFD sketch – gas recycle shown dotted since it is only needed if H_2 is used in (considerable) excess and must be recycled.



Route 1 is better since:

- Simpler PFD
- No gas recycle (no recycle compressor)
- No build up of inerts (CH_4) so recycle stream is not as large
- All products are valuable – fuel gas in Route 2 has a low value

12. a. Separate, purify, and recycle.

b. Recycle without separation and use a purge.

c. Recycle without separation and do not use a purge.

Acetaldehyde is a colorless liquid with a pungent, fruity odor. It is primarily used as a chemical intermediate, principally for the production of acetic acid, pyridine and pyridine bases, peracetic acid, pentaeythritol, butylene glycol, and chloral. Acetaldehyde is a volatile and flammable liquid that is miscible in water, alcohol, ether, benzene, gasoline, and other common organic solvents. In this problem, the synthesis of acetaldehyde via the dehydrogenation of ethanol is to be considered. The following reactions occur during the dehydrogenation of ethanol:

(1)

$$\underset{}{CH_3CH_2OH} \rightarrow \underset{\textit{acetaldehyde}}{CH_3CHO} + H_2$$

(2)

$$2CH_3CH_2OH \rightarrow \underset{\textit{ethyl acetate}}{CH_3COOC_2H_5} + 2H_2$$

(3)

13.

$$2CH_3CH_2OH \rightarrow \underset{\textit{butanol}}{CH_3(CH_2)_3OH} + H_2O$$

(4)

$$CH_3CH_2OH + H_2O \rightarrow \underset{\textit{acetic acid}}{CH_3COOH} + 2H_2$$

2.12 a. Good when product(s) and reactant(s) are easily separated and purified (most often by distillation.) Any inerts in the feed or byproducts can be removed by some unit operation and thus recycle does not require a purge.

b. When unused reactant(s) and product(s) are not easily separated (for example when both are low boiling point gases) and single pass conversion of reactant is low.

c. This is only possible when no significant inerts are present and any byproducts formed will react further or can reach equilibrium.

The conversion of ethanol is typically 60%. The yields for each reaction are approximately:

(1) acetaldehyde 92%

(2) ethyl acetate 4%

(3) butanol 2%

(4) acetic acid 2%

a. For this process, generate a process concept diagram showing all the input and output chemicals.

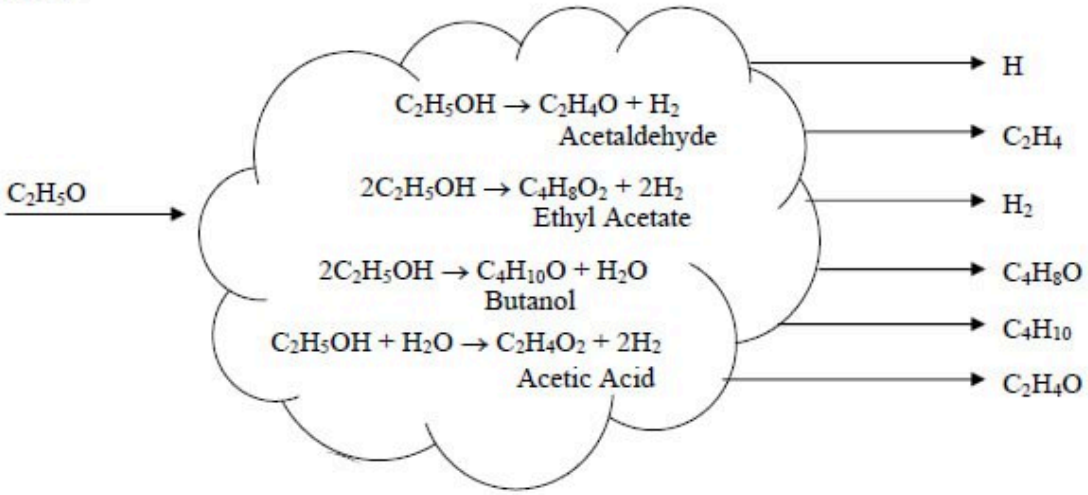
b. Develop two alternative preliminary process flow diagrams for this process.

Consider the following process in which liquid feed material *A* (normal BP of 110°C) is reacted with gaseous feed material *G* to produce main product *C* and by-products *R* and *S* via the following reactions:

$A + G \rightarrow C + S$

$G + C \rightarrow R$

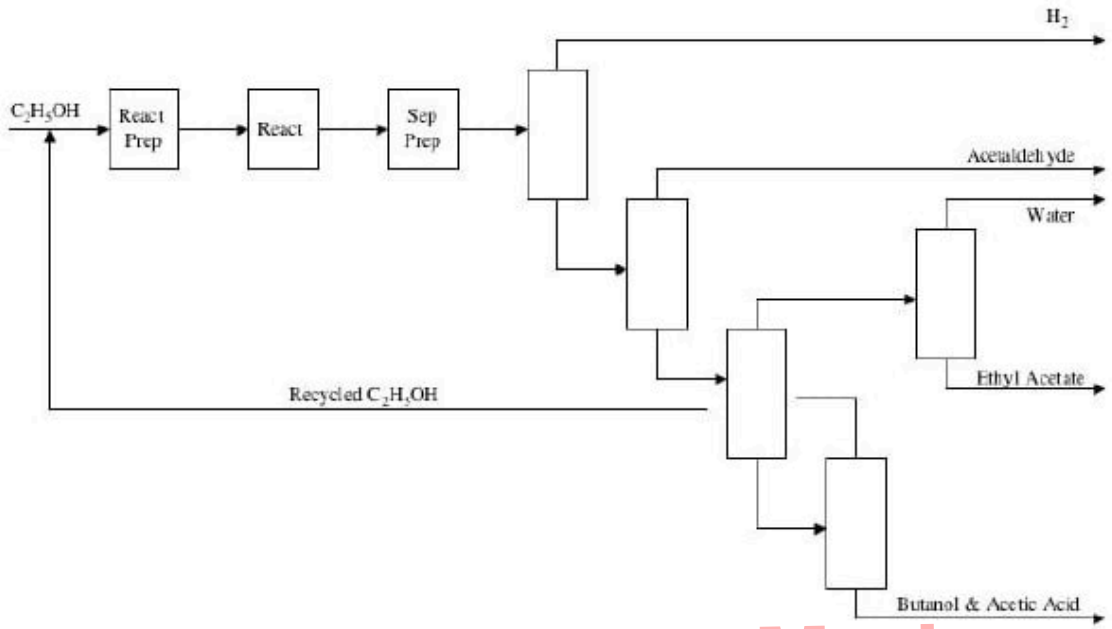
2.13 a.



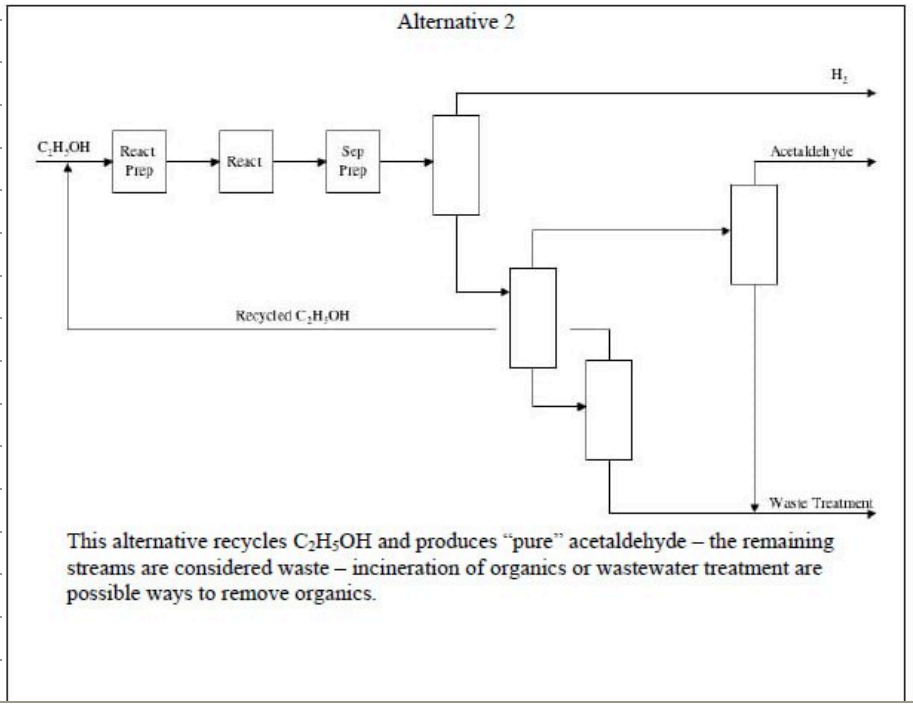
Order of volatility is acetaldehyde, water, ethyl acetate, ethanol, isobutanol, acetic acid.

b.

Alternative 1



Alternative 1 assumes butanol and acetic acid can be sold as a mixed product ⇒ very unlikely so probably have to add another column to separate.



Both feeds enter the process at ambient temperature and pressure. Both reactions occur in the gas phase at moderate temperature and pressure (250°C and 10 bar). The normal boiling points of G, S, and C are less than –120°C. By-product R has a normal boiling point of 75°C and is highly soluble inwater. Product C is very soluble in water but G and S are insoluble. The single-pass conversion through the reactor is low for feed A, and the ratio of G to A in the feed to the reactor should be maintained in excess of 4 to minimize the chance of other unwanted side reactions. Using this information, and assuming that both A and G are expensive, do the following.

a. Draw a preliminary process flow diagram identifying the main unit operations (reactors, compressors, pumps, heat exchangers, and separators), and identify the recycle structure of the process.

b. Justify the methods used to recycle A and G.

c. What unit operations do you suggest for your separators? Justify your choices.

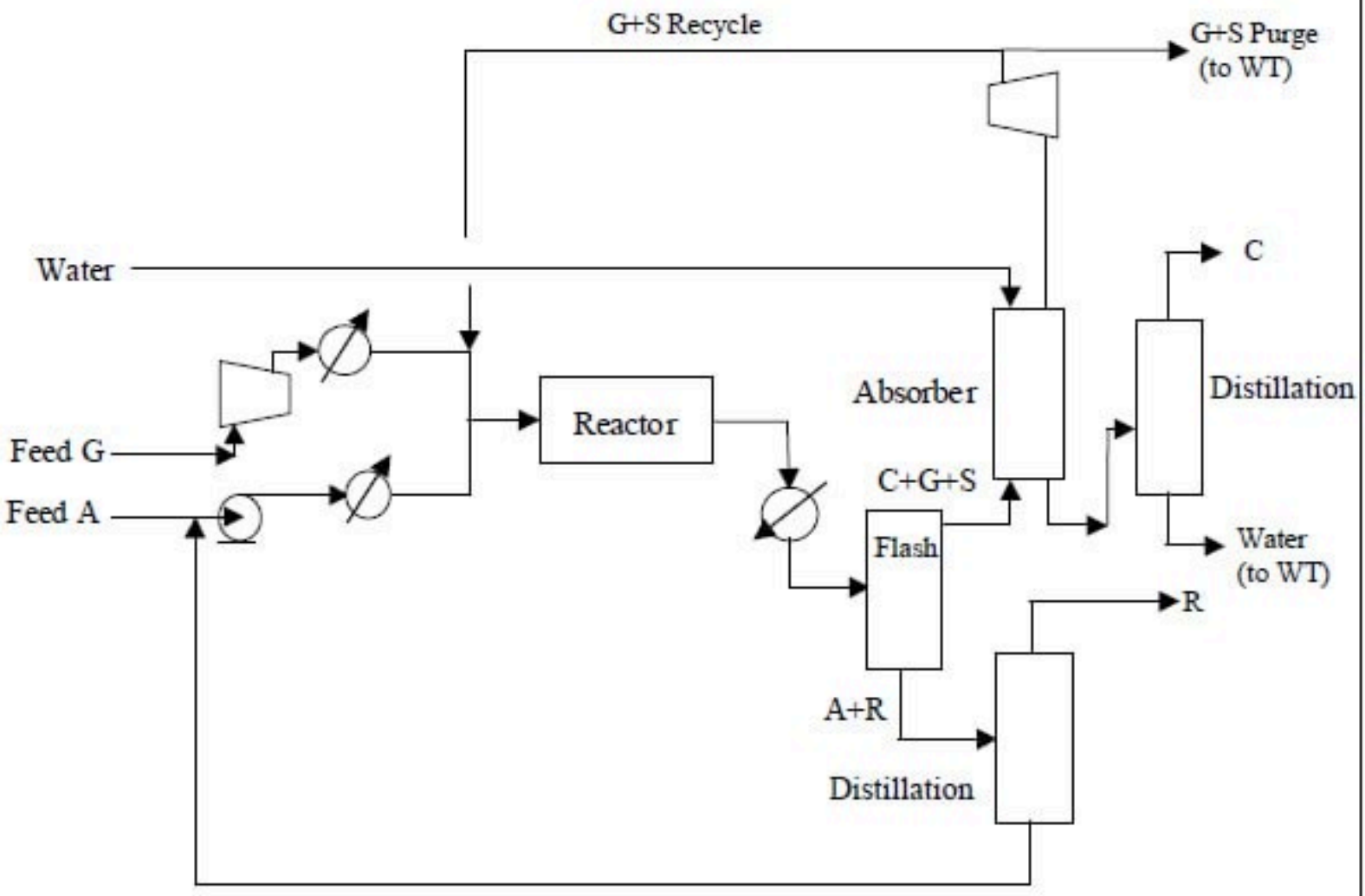
d. How would your PFD change if the price of feed material G were very low?

How is Scotch whisky made ?

2.14

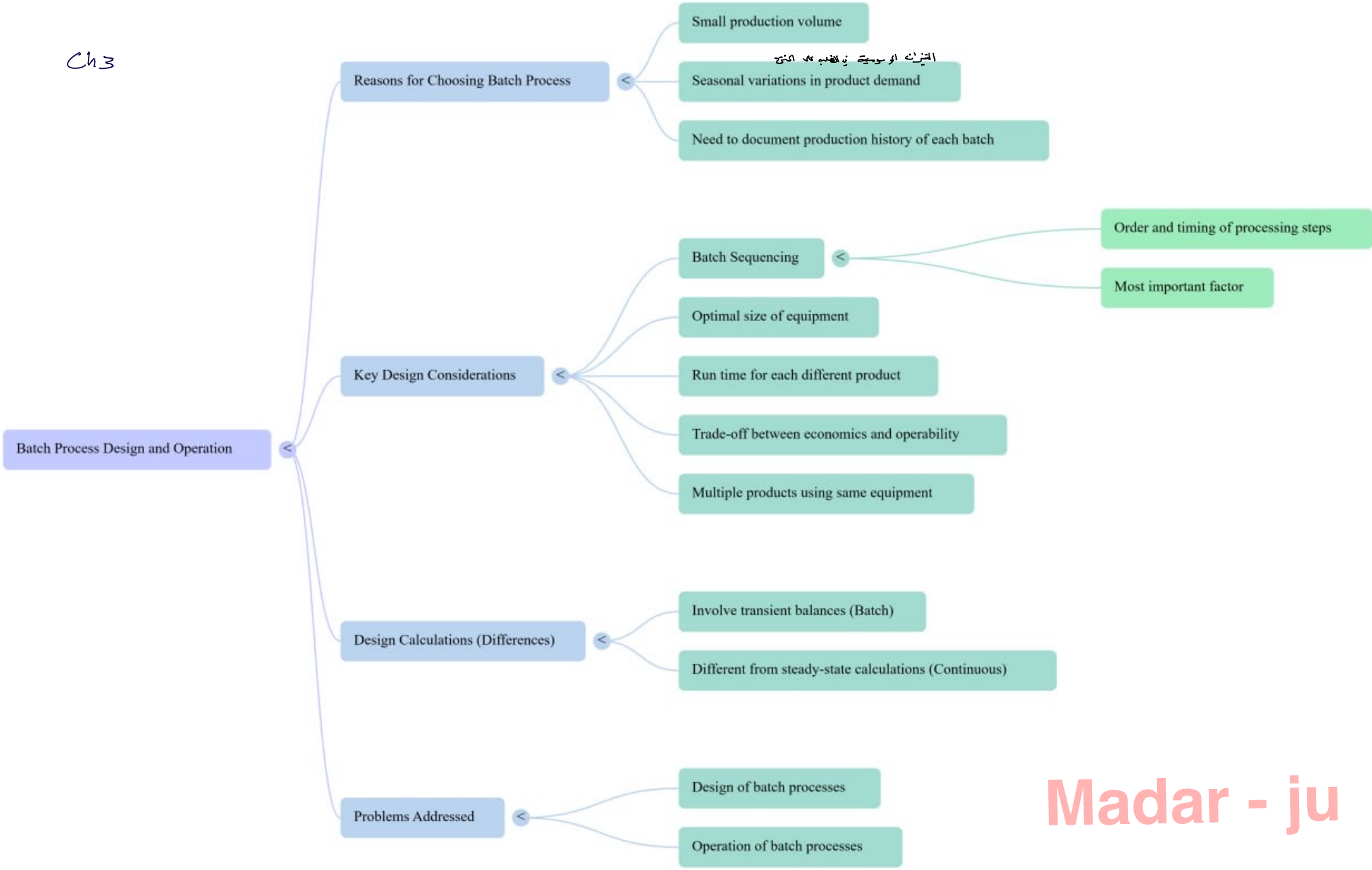
- A and R are both condensable and may be separated via distillation
- C may be separated by absorption into water
- R will be absorbed into water
- G and S cannot be separated except at very high pressure or low temperature

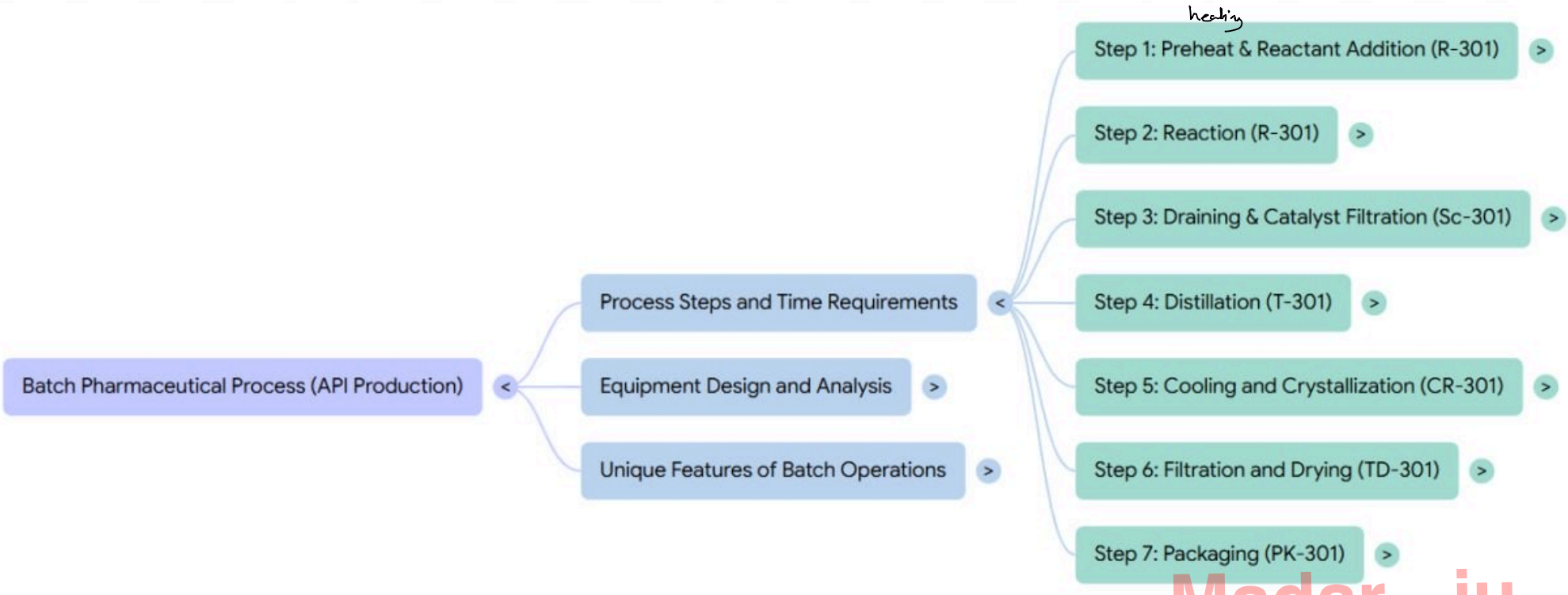
- After reaction, cool and condense A and R from other components.
- Separate A from R using distillation and recycle purified liquid A to the front end of the process
- Treat remaining gas stream in a water absorber to remove product C
- Separate C and from water via distillation
- Recycle unused G containing S – since S does not react further – we must add a purge to prevent accumulation of S in the system. This stream must be recycled as a gas using a recycle gas compressor.

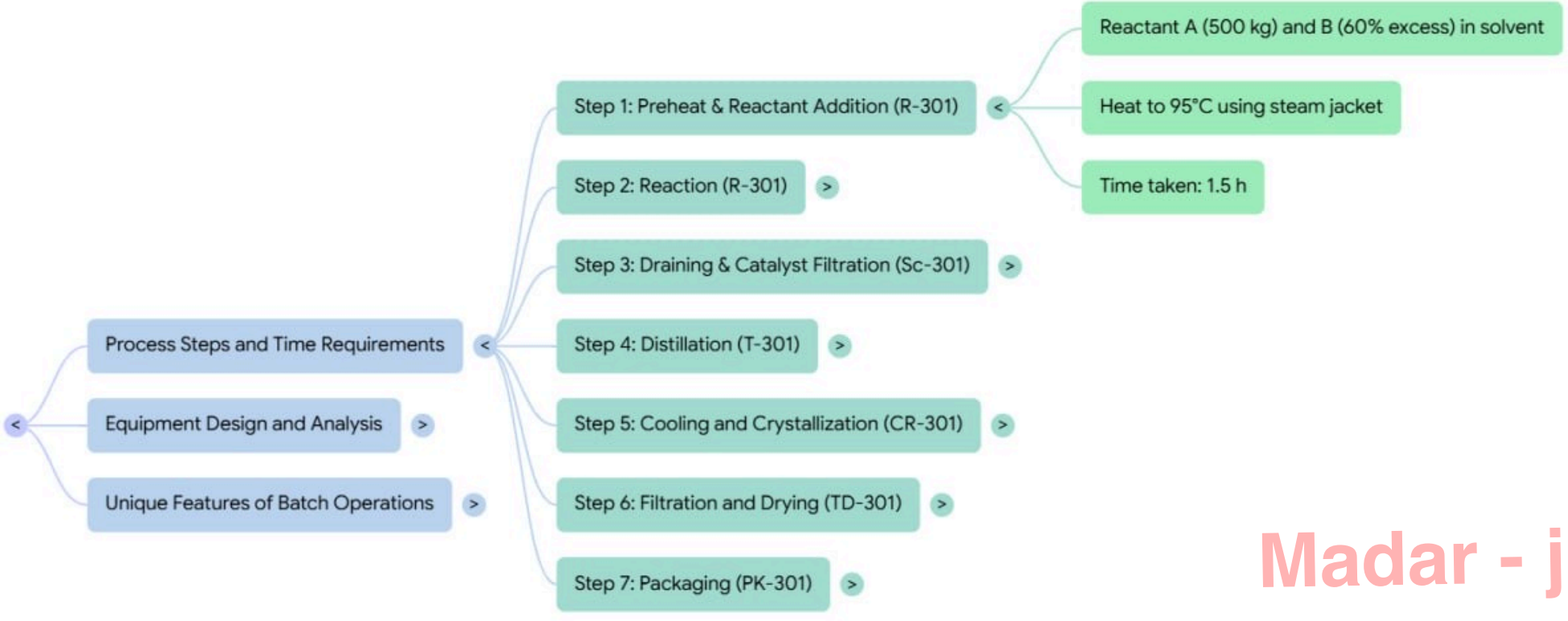


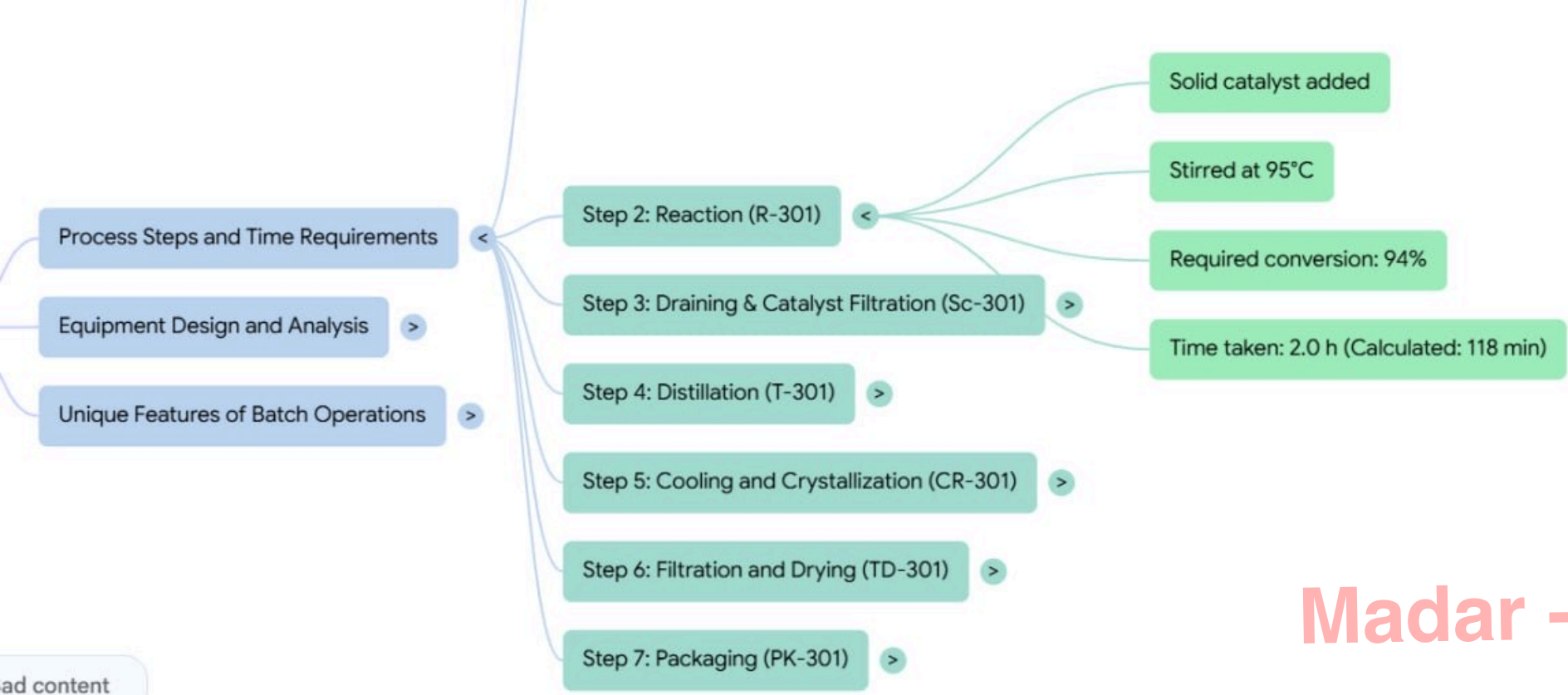
If the value of G was very low, then consider not recycling G (and S.)

Ch 3


Batch Process Design and Operation
Reasons for Choosing Batch Process
Small production volume
Seasonal variations in product demand
Need to document production history of each batch
Key Design Considerations
Batch Sequencing
Order and timing of processing steps
Most important factor
Optimal size of equipment
Run time for each different product
Trade-off between economics and operability
Multiple products using same equipment
Design Calculations (Differences)
Involve transient balances (Batch)
Different from steady-state calculations (Continuous)
Problems Addressed
Design of batch processes
Operation of batch processes


heating
Batch Pharmaceutical Process (API Production)
Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Step 1: Preheat & Reactant Addition (R-301)
Step 2: Reaction (R-301)
Step 3: Draining & Catalyst Filtration (Sc-301)
Step 4: Distillation (T-301)
Step 5: Cooling and Crystallization (CR-301)
Step 6: Filtration and Drying (TD-301)
Step 7: Packaging (PK-301)


Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Step 1: Preheat & Reactant Addition (R-301)
Step 2: Reaction (R-301)
Step 3: Draining & Catalyst Filtration (Sc-301)
Step 4: Distillation (T-301)
Step 5: Cooling and Crystallization (CR-301)
Step 6: Filtration and Drying (TD-301)
Step 7: Packaging (PK-301)
Reactant A (500 kg) and B (60% excess) in solvent
Heat to 95°C using steam jacket
Time taken: 1.5 h




Time taken: 2.0 h (Calculated: 118 min)
Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Step 3: Draining & Catalyst Filtration (Sc-301)
Mixture drained from reactor
Filter screen removes catalyst (stops reaction)
Time taken: 0.5 h (Calculated: 13 min, rounded to 30 min)
Step 4: Distillation (T-301)
Step 5: Cooling and Crystallization (CR-301)
Step 6: Filtration and Drying (TD-301)
Step 7: Packaging (PK-301)

rations >

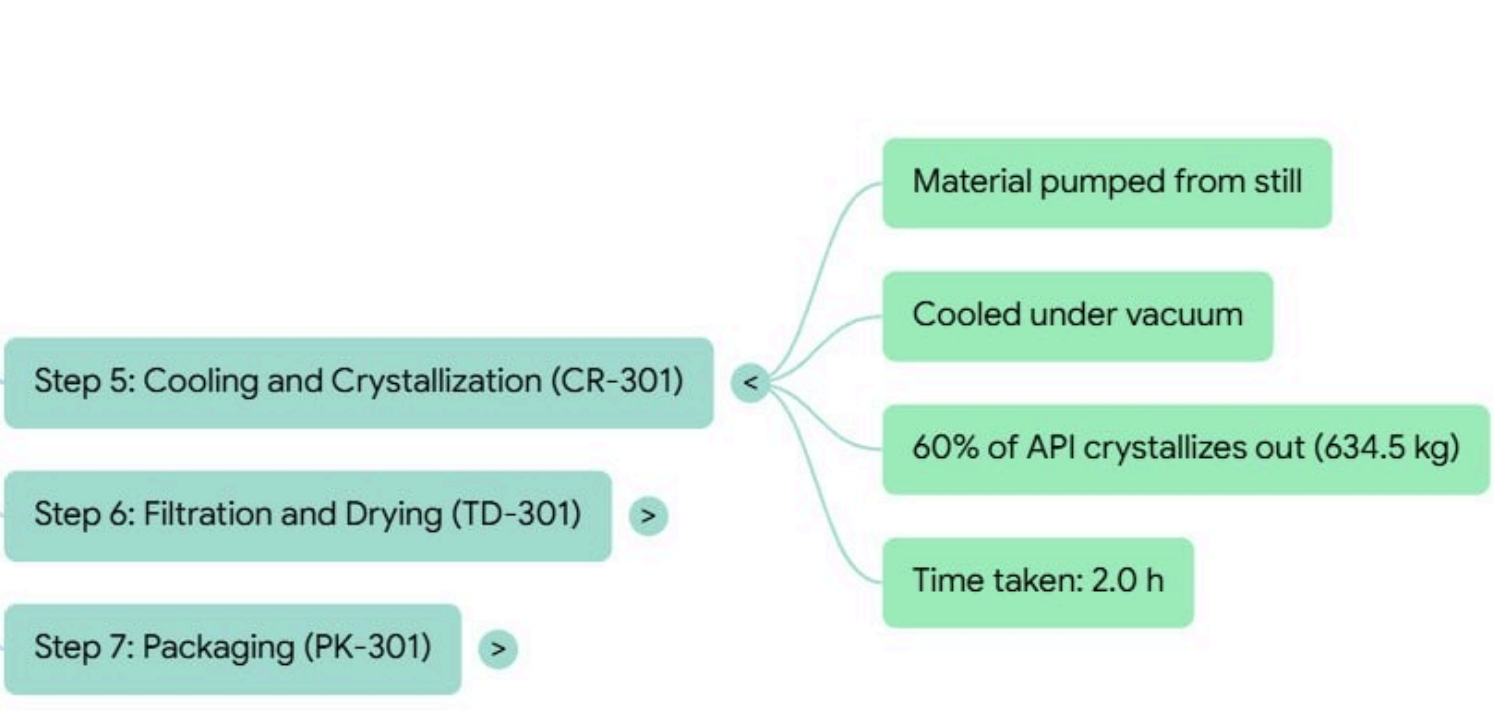
Step 4: Distillation (T-301) <

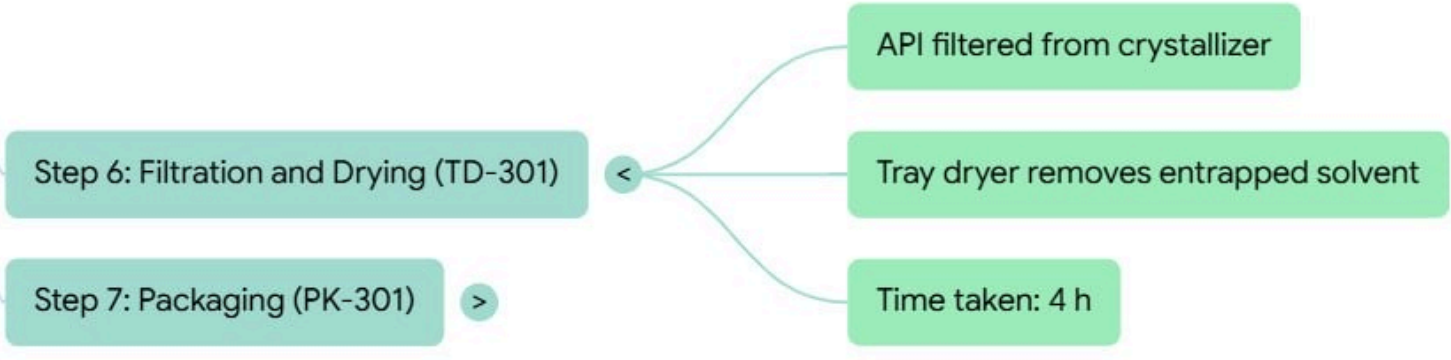
- Transferred to distillation column
- Distilled under vacuum
- Overhead product: Unused reactants and 50% solvent
- End-point: Still contains < 1 mol% Reactant B
- Total time: 3.5 h (Heating time: 70.3 min + Distillation time: 2.3 h)

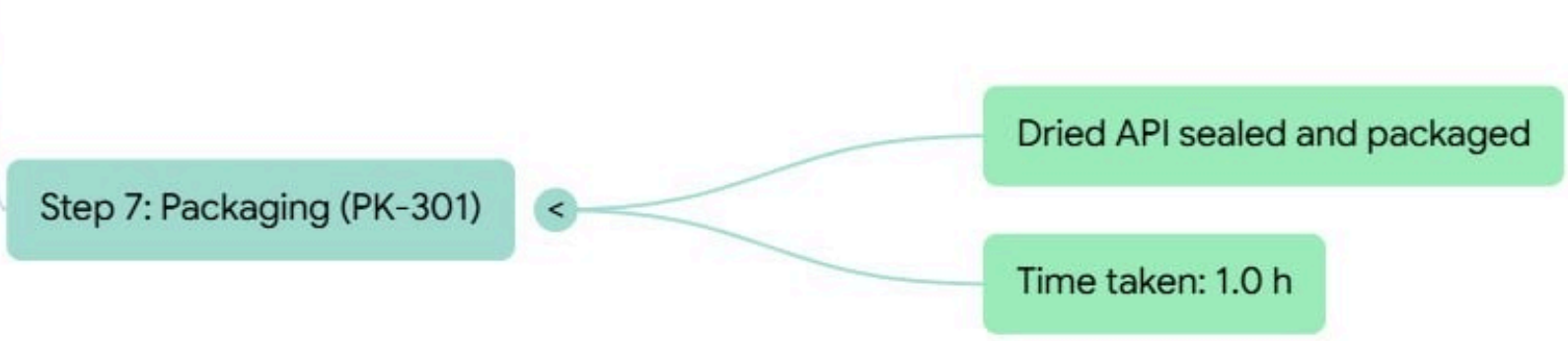
Step 5: Cooling and Crystallization (CR-301) >

Step 6: Filtration and Drying (TD-301) >

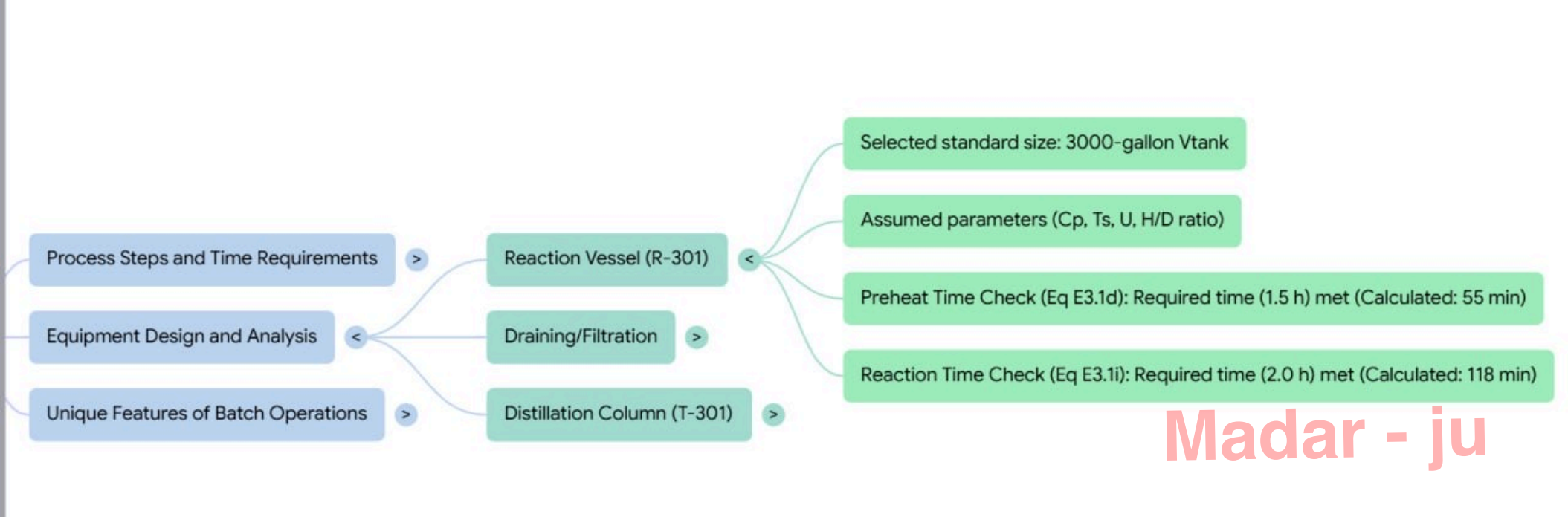
Step 7: Packaging (PK-301) >

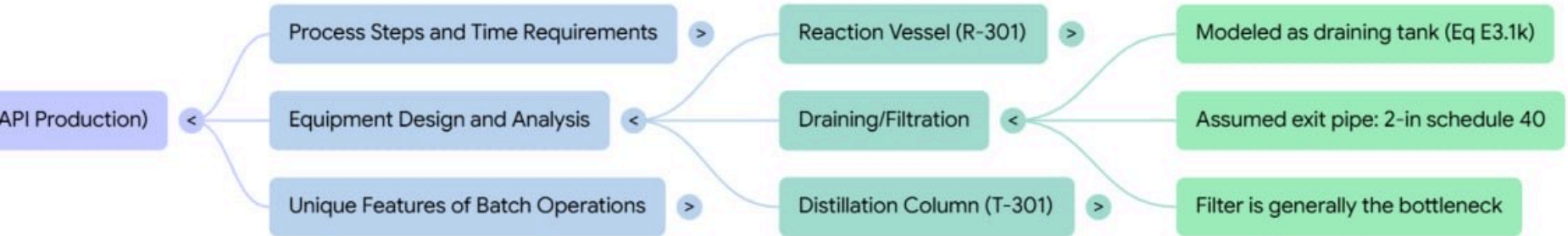

Step 5: Cooling and Crystallization (CR-301)
Material pumped from still
Cooled under vacuum
60% of API crystallizes out (634.5 kg)
Time taken: 2.0 h
Step 6: Filtration and Drying (TD-301)
Step 7: Packaging (PK-301)

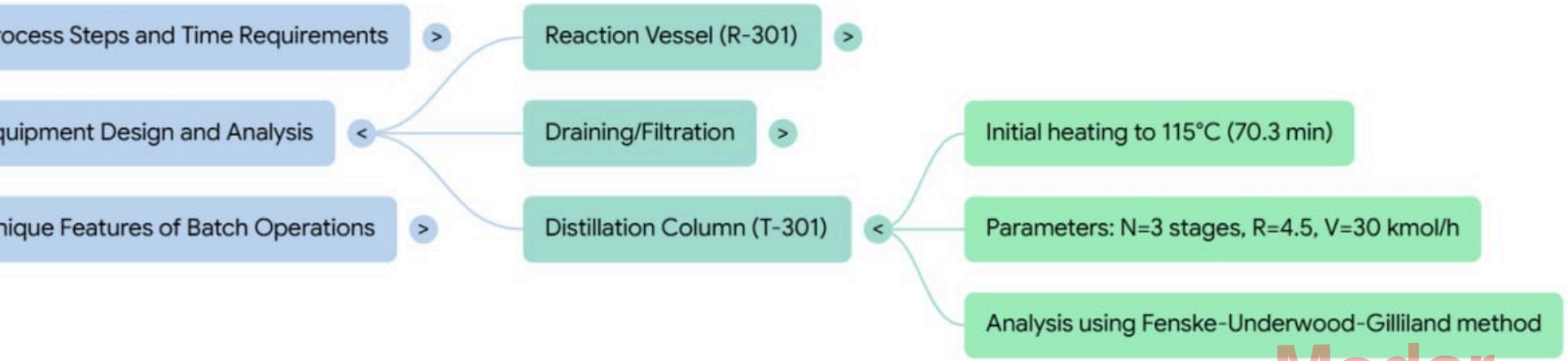

Step 6: Filtration and Drying (TD-301)
<
API filtered from crystallizer
Tray dryer removes entrapped solvent
Time taken: 4 h
Step 7: Packaging (PK-301)
>

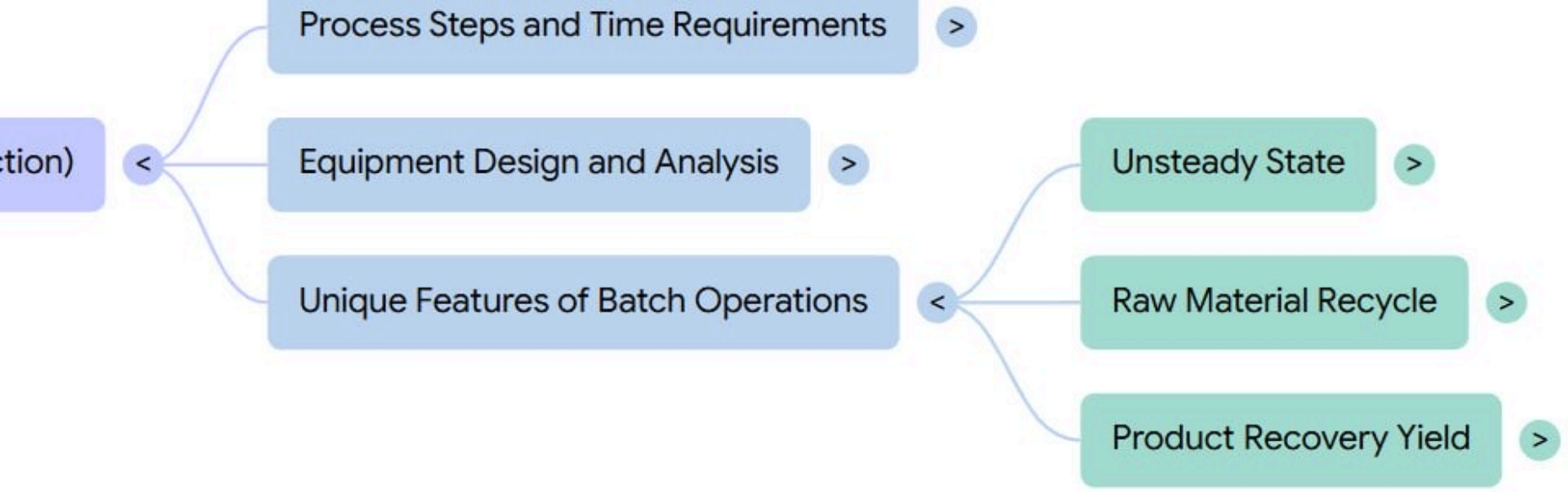

Step 7: Packaging (PK-301)
<
Dried API sealed and packaged
Time taken: 1.0 h

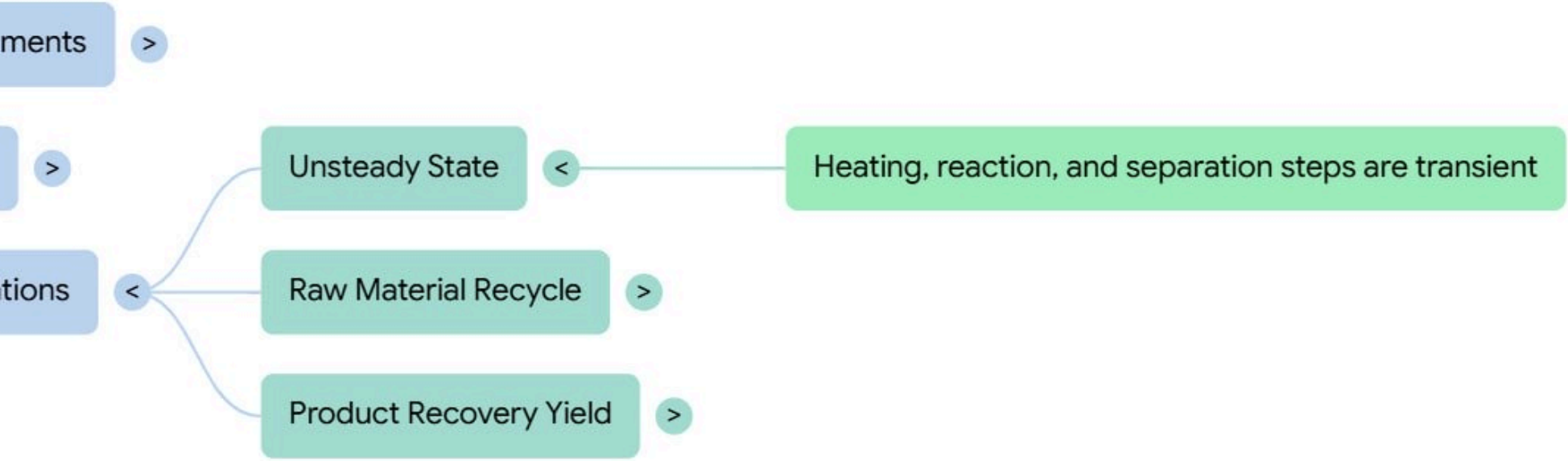

Batch Pharmaceutical Process (API Production)
Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Reaction Vessel (R-301)
Draining/Filtration
Distillation Column (T-301)


Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Reaction Vessel (R-301)
Draining/Filtration
Distillation Column (T-301)
Selected standard size: 3000-gallon Vtank
Assumed parameters (Cp, Ts, U, H/D ratio)
Preheat Time Check (Eq E3.1d): Required time (1.5 h) met (Calculated: 55 min)
Reaction Time Check (Eq E3.1i): Required time (2.0 h) met (Calculated: 118 min)


API Production)
Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Reaction Vessel (R-301)
Draining/Filtration
Distillation Column (T-301)
Modeled as draining tank (Eq E3.1k)
Assumed exit pipe: 2-in schedule 40
Filter is generally the bottleneck


rocess Steps and Time Requirements
quipment Design and Analysis
nique Features of Batch Operations
Reaction Vessel (R-301)
Draining/Filtration
Distillation Column (T-301)
Initial heating to 115°C (70.3 min)
Parameters: N=3 stages, R=4.5, V=30 kmol/h
Analysis using Fenske-Underwood-Gilliland method


ction)
Process Steps and Time Requirements
Equipment Design and Analysis
Unique Features of Batch Operations
Unsteady State
Raw Material Recycle
Product Recovery Yield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Unsteady State
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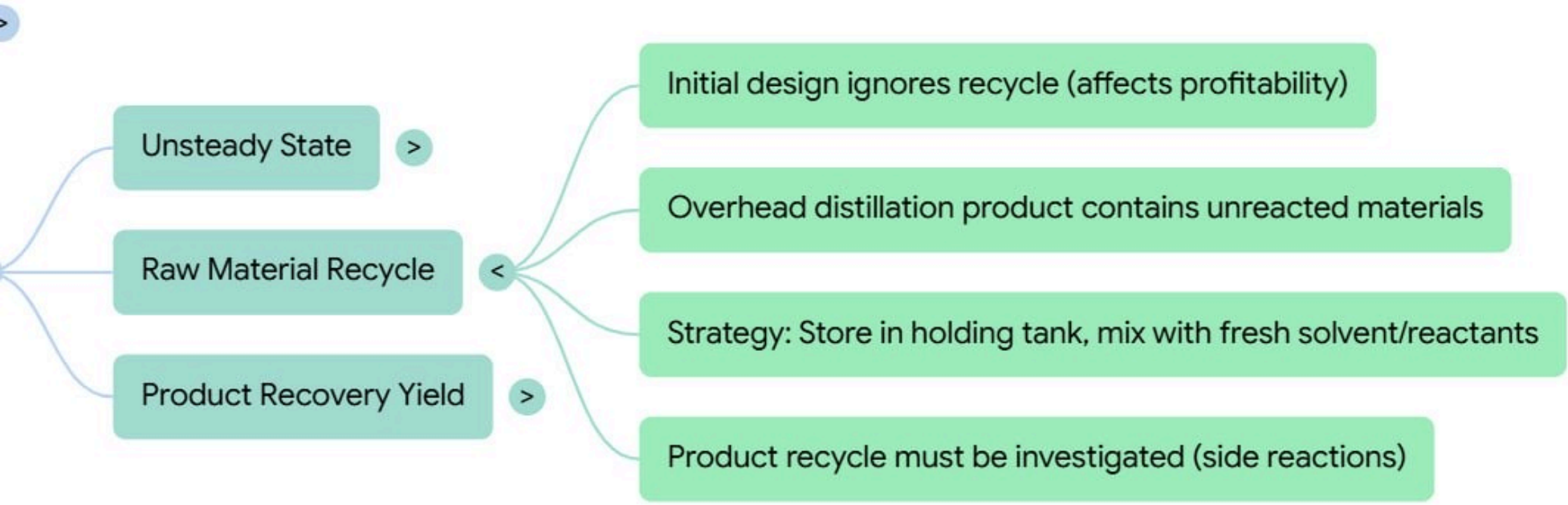
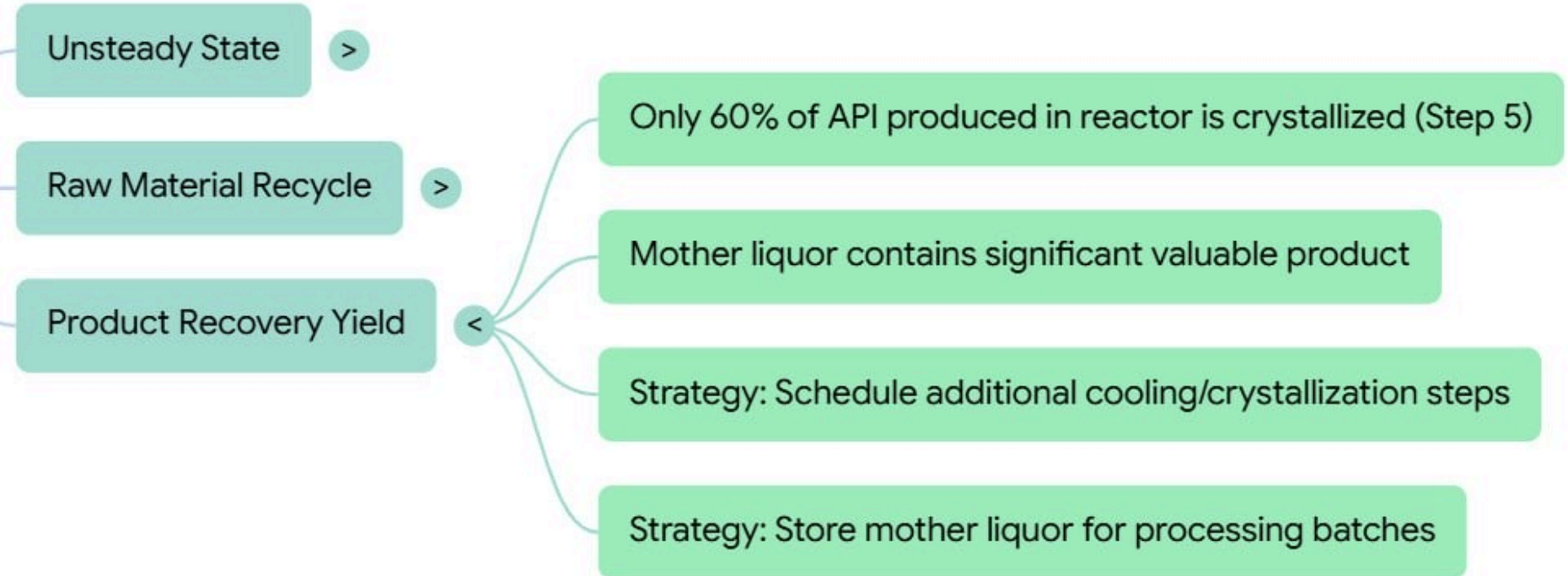
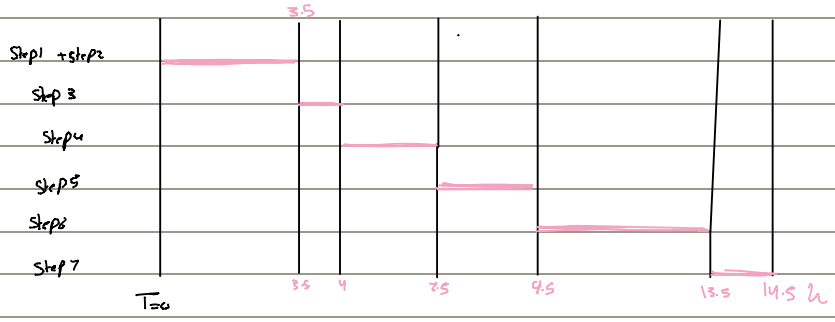
Heating, reaction, and separation steps are transient
Raw Material Recycle
>
Product Recovery Yield
>


Unsteady State
Raw Material Recycle
Product Recovery Yield
Initial design ignores recycle (affects profitability)
Overhead distillation product contains unreacted materials
Strategy: Store in holding tank, mix with fresh solvent/reactants
Product recycle must be investigated (side reactions)


Unsteady State
>
Raw Material Recycle
>
Product Recovery Yield
<
Only 60% of API produced in reactor is crystallized (Step 5)
Mother liquor contains significant valuable product
Strategy: Schedule additional cooling/crystallization steps
Strategy: Store mother liquor for processing batches

Gantt charts and scheduling



5000 / 634.5 kg = 8

Non overlapping $T_{No} = n \sum t_i$

$= 8 \times (3.5 + 0.5 + 3.5 + 2 + 4 + 1) = 118\,h$

$T_o = T = (n-1) \max(t_i) + \sum t_i$

overlapping $= (8-1) \times 4 + 14.5 = 42.5\,h$

$t_{cycle\,No} = \sum t_i$

$t_{cycle\,o} = \frac{T}{n} = \frac{(n-1)\max(t_i) + \sum t_i}{n}$ exact

$t_{cycle} = \max t_i$ approximated

Example 3.3

$T_{product} = (X-1) \max t_i + \sum t_i$

$500 = ((X-1)(2.5) + 8) + ((X-1)(4.5) + 9.5) + ((X-1)(4.5) + 11)$

$X = 42$ batch

$t_{cycle\,A} = \frac{(42-1)(2.5) + 8}{42} = 2.631$

$t_{cycle\,B} = \frac{(42-1)(4.5) + 9.5}{42} = 4.619$

$t_{cycle\,C} = \frac{(42-1)(4.5) + 11}{42} = 4.855$

Flowshop

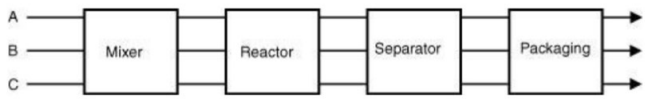
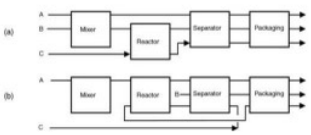


Figure 3.5 Two Examples of Jobshop Plants for Three Products A, B, and C



Chapter 3

3.1. What is a flowshop plant?
A flowshop plant is a plant in which several batch products are produced using all or a subset of the same equipment and in which the operations for each batch follow the same sequence. Thus the flow of any batch through equipment A, B, C, D … is always A→B→C→D→…. Omissions of equipment are possible but no reversal in direction is allowed.

3.2. What is a jobshop plant?
A flowshop plant is a plant in which several batch products are produced using all or a subset of the same equipment but for which the operations of at least one batch product do not follow the same sequence, e.g., A→C→D→B

3.3. What are the two main methods for sequencing multiproduct processes?
Either use multi-product campaigns or multiple single-product campaigns.

3.4. Give one advantage and one disadvantage of using single-product campaigns in a multiproduct plant.

Advantage – sequencing of single-product campaigns is relatively simple and repetitive and probably less prone to operator error since the batch recipe remains the same over the entire campaign.

Disadvantage – significant final product storage will be required since all products will not be made all the time and in order to even out supply some inventory of products will have to be maintained in storage. Single-product campaigns may be less efficient than multi-product campaigns.

3.5. What is the difference between a zero-wait and a uis process?

material moves immediately from one unit to the next
A zero-wait process is one in which the batch is transferred immediately from the current piece of equipment to the next piece of equipment in the recipe sequence. This type of process eliminates the need for intermediate storage (storage of unfinished products or intermediates).

can be stored in any amount
A uis (unlimited intermediate storage) process is one in which any amount of any intermediate product may be stored. Such a process maximizes the use of the processing equipment but obviously requires an unlimited amount of storage.

more flexible

3.6 Number of batches of A is twice that for B or C – repeat Example 3.3 with this restriction using a 500 h cycle time.

Table E3.3: Equipment times needed to produce A, B, and C

Product	Time in Mixer	Time in Reactor	Time in Separator	Time in Packaging
A	1.5	1.5	2.5	2.5
B	1.0	2.5	4.5	1.5
C	1.0	4.5	3.5	2.0

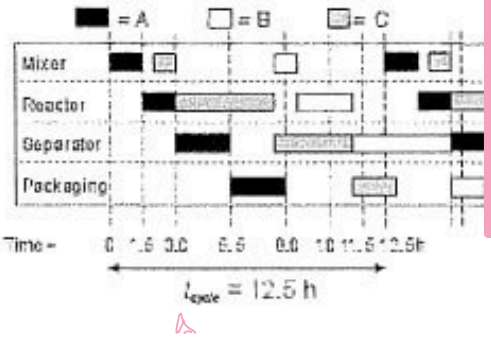
Using Equation (3.6) with $t_{cycle,A} = 2.5$, $t_{cycle,B} = 4.5$, and $t_{cycle,C} = 4.5$

If x is the number of batches of Products B and C, then 2x is the number of batches of Product A

$$T = 500 = 2x(2.5) + x(4.5 + 4.5) \Rightarrow x = \frac{500}{14} = 35.7$$

Number of batches for each product are A = 70, B=35, C=35

3.7 For Examples 3.3 and 3.4, determine the number of batches that can be produced in a month (500 h) using a multi-product campaign strategy with the sequence ACBACBACB. Are there any other sequences for this problem other than the one used in Example 3.4 and the one used here?



The multi-product cycle time = 2.5 + 2.0 + 3.5 + 4.5 = 12.5 h

Number of batches per month = (500)/(12.5) = 40 each of A, B, and C

The only sequences that can be used for multi-product campaigns are ABCABCABC (Example 3.4) and ACBACBACB as used here.

3.8 Consider the multi-product batch plant described in Table P3.8

Table P3.8: Equipment Processing Times for Processes A, B, and C

Process	Mixer	Reactor	Separator
A	2.0 h	5.0 h	4.0 h
B	3.0 h	4.0 h	3.5 h
C	1.0 h	3.0 h	4.5 h

It is required to produce the same number of batches of each product. Determine the number of batches that can be produced in a 500 h operating period using the following strategies:

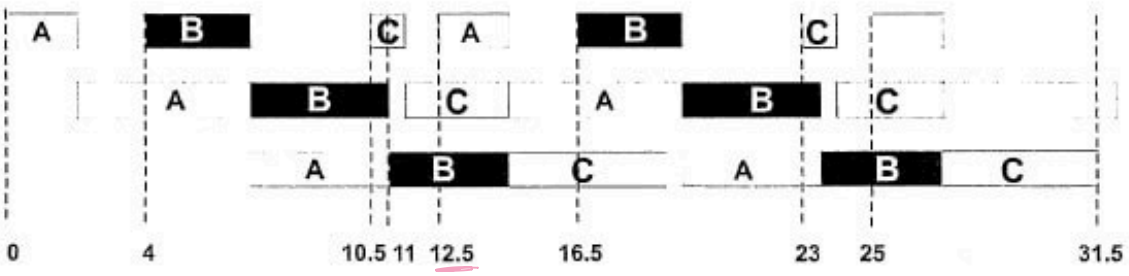
(a) using single-product campaigns for each product

Using Equation (3.6) with $t_{cycle,A} = 5.0$, $t_{cycle,B} = 4.0$, and $t_{cycle,C} = 4.5$

$$T = 500 = x(5.0+4.0+4.5) \Rightarrow x = \frac{500}{13.5} = 37.0$$

$x = 37$ batches

(b) using a multi-product campaign using the sequence ABCABCABC...



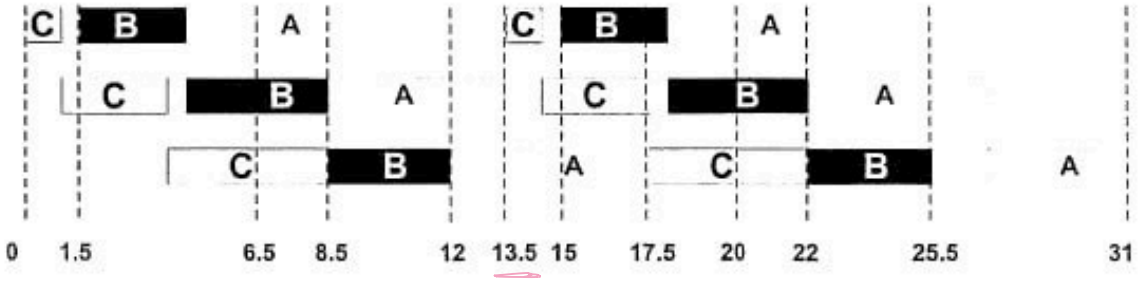
From this diagram we see that the cycle time for the multi-product campaign using the sequence ABC is 12.5 h.

Therefore, the number of batches, x, of each product that can be made during a 500 h period is given by:

$$T = 500 = 12.5x \Rightarrow x = \frac{500}{12.5} = 40$$

$x = 40$ batches

(c) using a multi-product campaign using the sequence CBACBACBA…



From this diagram we see that the cycle time for the multi-product campaign using the sequence ABC is 13.5 h.
Therefore, the number of batches, x, of product that can be made during a 500 h period is given by:

$$T = 500 = 13.5x \Rightarrow x = \frac{500}{13.5} = 37.0$$

$x = 37$ batches

3.9 Consider the process given in Problem 3.8. Assuming that a single-product campaign strategy is repeated every 500 h operating period and further assuming that the production rate (for a year = 6,000 h) for products A, B, C are 18,000 kg/y, 24,000 kg/y, and 30,000 kg/y, respectively, determine the minimum volume of product storage required. Assume that the product densities of A, B, and C are 1100, 1200, and 1000 kg/m^3, respectively

The tables below shows the results using data given from Problem 8

Rate	**Product A**	**Product B**	**Product C**
Volume (m^3) of product required per month	18,000/12/1,100 = 1.36	24,000/12/1,200 = 1.67	30,000/12/1,000 = 2.5
Cycle time (h)	5.0	4.0	4.5
Production rate, r_p (m^3/h)	(1.36)/(37)(5) = 0.007371	(1.67)/(37)(4) = 0.01126	(2.5)/(37)(4.5) = 0.015015
Demand rate, r_d (m^3/h)	(1.36)/(500) = 0.002727	0.003333	0.005

Product	Campaign time, t_{camp} (h)	r_p-r_d (m^3/h)	Minimum volume of product storage, V_s (m^3)
A	(37)(5) = 185	0.007371 – 0.00273 = 0.004644	(0.004644)(185) = **0.859**
B	(37)(4) = 148	0.01126 – 0.003333 = 0.007928	(0.007928)(148) = **1.173**
C	(37)(4.5) = 166.5	0.015015 – 0.005 = 0.010015	(0.010015)(166.5) = **1.668**

3.10

Table P3.10A: Production rates for A, B, and C

Product	Yearly production	Production in 500 h
A	150,000 kg	12,500 kg
B	210,000 kg	17,500 kg
C	360,000 kg	30,000 kg

Table P3.10B: Specific Reactor/Mixer Volumes for Processes A, B, and C

Process	A	B	C
v_{react} (m^3/kg-product)	0.0073	0.0095	0.0047
t_{cycle} (h)	6.0	9.5	18.5

Let the single-product campaign times for the three products be t_A, t_B, and t_C, respectively.

Applying Equation (3.6), the following relationship is obtained:

$$t_A + t_B + t_C = 500 \tag{3.9}$$

The number of campaigns per product is then given by $t_x/t_{cycle,x}$ and

$$\text{batch size (kg/batch)} = \frac{\text{production of x}}{t_x / t_{cycle,x}} \tag{3.10}$$

Furthermore, the volume of a batch is found by multiplying Equation (3.10) by $v_{reac,x}$, and equating batch volumes for the different products yields:

$$\text{Volume of batch} = \frac{(\text{production of x})(v_{react,x})}{t_x / t_{cycle,x}} \tag{3.11}$$

$$\frac{(12{,}500)(.0073)}{t_A/6.0} = \frac{(17{,}500)(.0095)}{t_B/9.5} = \frac{(30{,}000)(.0047)}{t_C/18.5} \tag{3.12}$$

$\frac{547.5}{t_A} = \frac{1579.375}{t_B} = \frac{2608.5}{t_C}$

$t_B = 2.8847\ t_A$

$t_C = 4.76\ t_A$

$t_A + t_B + t_C = 500$

$t_A = 57.8$

Solving Equations (3.9) and (3.12), yields:

$$t_A = 57.8\ h$$
$$t_B = 166.8\ h$$
$$t_C = 275.4 h$$
$$v_{react,A} = v_{react,B} = v_{react,C} = 9.47\ m^3$$

#batches per campaign for product A = t_A / 6.0 = 9.6
#batches per campaign for product B = t_B / 9.5 = 17.6
#batches per campaign for product C = t_C / 18.5 = 14.9

Clearly the number of batches should be an integer value. Rounding these numbers yields

For product A
Number of batches = 10
t_A = (10)(6.0) = 60 h
V_A = (12,500)(0.0073)/(10) = 9.13 m^3

For product B
Number of batches = 17
t_B = (17)(9.5) = 161.5 h
V_B = (17,500)(0.0095)/(17) = 9.78 m^3

For product C
Number of batches = 15
t_C = (15)(18.5) = 277.5 h
V_C = (30,000)(0.0047)/(15) = 9.40 m^3

Total time for production cycle = 499 h ~ 500 h
Volume of reactor = 9.78 m^3 (limiting condition for Product B)

Chapter 11 Summary
Utilizing Experience-Based Principles to Confirm the Suitability of a Process Design

Chapter Summary

- This chapter emphasizes the use of engineering experience and heuristics to evaluate whether a process design is realistic and practical.
- A design that satisfies mass and energy balances may still be unsafe, uneconomical, or impossible to construct.
- Experience-based principles are used as sanity checks to validate simulation and design results.
- These principles help identify design flaws early, reducing the risk of costly redesigns later.
- They do not replace detailed calculations but complement them.
- The chapter focuses on engineering judgment rather than numerical optimization.

Key Experience-Based Checks

- Temperature and Pressure: Operating conditions should be within reasonable industrial limits.
- Equipment Size: Reactors, columns, and heat exchangers must be physically and mechanically feasible.
- Process Flow Logic: Unit operations should be arranged in a logical and conventional sequence.
- Energy Usage: Utility consumption should be reasonable, and heat integration opportunities should be considered.
- Safety: High temperatures, high pressures, and hazardous materials must be carefully evaluated.
- Economic Feasibility: Excessive complexity usually leads to high capital and operating costs.

Expected Exam Questions

- **Q1.** Explain the purpose of using experience-based principles in process design.
- **Q2.** Why is satisfying mass and energy balances not sufficient to ensure a feasible process design?
- **Q3.** Define experience-based principles (heuristics) and give two examples.
- **Q4.** List four checks an engineer should perform to confirm the suitability of a process design.
- **Q5.** A distillation column requires 400 stages and a diameter of 20 m. What experience-based concerns does this raise?
- **Q6.** True or False: If a process simulation converges, the design is automatically practical. Justify your answer.

Key Sentence to Remember for Exams

Experience-based principles are used to verify that a process design is realistic, safe, and economically reasonable, and to identify potential problems that are not revealed by calculations alone.

Chapter 11 – Questions & Answers
Utilizing Experience-Based Principles to Confirm the Suitability of a Process Design

Q1. What is the main purpose of Chapter 11?

Answer: The main purpose of Chapter 11 is to show how experience-based principles (heuristics) are used to check whether a process design is realistic, practical, safe, and economically feasible. The chapter emphasizes engineering judgment rather than detailed calculations.

Q2. Why is satisfying mass and energy balances not enough to ensure a feasible design?

Answer: Because a design may satisfy mass and energy balances but still be impractical, unsafe, too expensive, or impossible to construct. Experience-based checks help identify such problems early.

Q3. What are experience-based principles (heuristics)?

Answer: Experience-based principles are rules of thumb developed from industrial experience. They are used as sanity checks to evaluate the suitability of a process design and to validate simulation results.

Q4. Give four examples of experience-based checks used in process design.

Answer: Examples include: (1) checking that operating temperatures and pressures are within reasonable industrial limits, (2) verifying that equipment sizes and numbers are practical, (3) ensuring the logical order of unit operations, and (4) evaluating energy consumption and safety issues.

Q5. Why are temperature and pressure levels important in experience-based evaluation?

Answer: Extreme temperatures or pressures often require special materials, increase costs, and raise safety concerns. Experience-based principles help engineers avoid unrealistic operating conditions.

Q6. A distillation column requires 400 stages and a diameter of 20 m. What concerns arise?

Answer: Such a column is mechanically impractical and extremely expensive. This indicates a poor separation choice or unrealistic design assumptions.

Q7. How does Chapter 11 relate to process simulation?

Answer: Chapter 11 emphasizes that simulation results must be checked using engineering judgment. A converged simulation does not guarantee a practical or safe design.

Q8. What role does safety play in confirming process suitability?

Answer: Safety is critical. High pressures, high temperatures, and hazardous materials must be carefully reviewed to ensure that the process can be operated safely.

Q9. How do experience-based principles affect economics?

Answer: They help identify overly complex designs, excessive equipment, and extreme operating conditions, all of which increase capital and operating costs.

Q10. True or False: If a process simulation converges, the design is automatically practical.

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Answer: False. A converged simulation may still represent an impractical, unsafe, or uneconomical process design.

Key Exam Sentence

Experience-based principles are used to confirm that a process design is realistic, safe, and economically reasonable, and to detect problems that calculations alone may not reveal.

Ch11:

Example 11.1

T = 93 °C u' = 3.05 m/s D = 38 mm New condition

T = 21°C u = 1.83 m/s h = 5250 W/m².°C old condition

(11.1)

$$hD/k = (0.023)(Du\rho/\mu)^{0.8}(C_p\mu/k)^{1/3}$$

Property	21°C (70°F)	93°C (200°F)	Ratio of (New/Old)
ρ (kg/m³)	997.4	963.2	0.966
k (W/m°C)	0.604	0.678	1.12
C_p (kJ/kg°C)	4.19	4.20	1.00
μ (kg/m/s)	9.8×10^{-4}	3.06×10^{-4}	0.312

Step 1: Predict

h'_{pred} = 5250 W/m².°C

Step 2: Authenticate / Analyze

$Re = \frac{\rho u D}{\mu} = \frac{(997.4)(1.83)(0.0381)}{9.8 \times 10^{-4}} = 7.1 \times 10^4$ → turbulent.

Sieder – Tate ratio

$$\frac{h'}{h} = \left(\frac{D'}{D}\right)^{0.2}\left(\frac{u'}{u}\right)^{0.8}\left(\frac{\rho'}{\rho}\right)^{0.8}\left(\frac{\mu}{\mu'}\right)^{0.47}\left(\frac{cp'}{cp}\right)^{0.33}\left(\frac{k'}{k}\right)^{0.67}$$

velocity term · Density term · Viscosity · heat Capacity · Thermal Conductivity

h'/h = 2.725

h' = 14300 W/m².°C

at T = 98°C

h' = 1.43×10^4 W/m².°C

Example 11.2

a) V-102

L/D = 2.5 to 5 optimum = 3 L = 2.5 D = 2.5(1.33) = 3.3 m

liquid hold up time = 5 min

$u = K\sqrt{\frac{\rho_L}{\rho_v} - 1}$ K = 0.0305

$u_{act} = 0.75\ u$

$u = 0.0305\sqrt{\frac{950}{8} - 1} = 0.313$ m/s

$u_{act} = 0.75 \times 0.313 = 0.23$ m/s

$V_L = 0.5\left(L\frac{\pi D^2}{4}\right) = 0.726L$

$V_S = \frac{(5)(60)(11570)}{950.3/100} = 1.13\ m^3$

$0.726L = 1.13 \rightarrow L = 1.56$ m

$L/D = \frac{1.56}{1.33} = 1.17$

b) E-105

$F = 0.9$

$\Delta T_{min} = 10°C$

Cooling water $30°C \longrightarrow 40°C$

$U = 850\ W/m^2.°C$

$Q = 1085\ MJ/h = 301\ kW = 301000\ W$

hot stream $105°C \longrightarrow 38°C$

1) LMTD $\Delta T_{lm} = \frac{(105-40)-(38-30)}{\ln\left(\frac{105-40}{38-30}\right)} = 27.2°C$

2) Area $A = \frac{Q}{U \Delta T_{lm} F} = \frac{301000}{(850)(27.2)(0.9)} = 14.4\ m^2$

c) P-101

$\dot{m} = 13300\ kg/h$

$\rho = 870\ kg/m^3$

$\Delta P = 25.8 - 1.2 = 24.6\ bar$

$Flow = \frac{13300}{60 \cdot 870} = 0.255\ m^3/min$

$P_{P,ideal} = 11.8(0.255)(24.6) = 10.5\ kW$

$\epsilon = 0.75$

$P_{shaft} = \frac{10.5}{0.75} = 14\ kW$

d) C-101

$W_{ideal} = \dot{n} 2 R T_1 \frac{(P_2/P_1)^a - 1}{a}$, $W_{actual} = \frac{W_{ideal}}{\eta}$

$flow = 6770\ kg/h$, $MW = 8.45$

$T_1 = 38°C = 311\ K$

$P_1 = 23.9\ bar$, $P_2 = 25.8\ bar$

$k = 1.41 \longrightarrow a = 0.2908$

$\eta = 0.75$

$\dot{n} = \frac{6770}{3600 \cdot 8.45} = 0.223\ kmol/s$

$W_{ideal} = 37.7\ kW \longrightarrow W_{actual} = \frac{37.7}{0.75} = 50.3\ kW$

e) T-101

$x_{benz} = 0.9962$, $x_{tol} = 0.0038$

$\alpha_{benz} = 2.41$, $\alpha_{tol} = 2.13$

$\alpha_{avg} = \sqrt{(2.41)(2.13)} = 2.28$

$N_{min} = \frac{\ln\left(\frac{x/(1-x)_{benz}}{x/(1-x)_{tol}}\right)}{\ln(\alpha_{avg})} = 10.9$

$R_{min} = \frac{F/D}{\alpha - 1} = 1.05$

$R = (1.2 \text{ to } 1.5) R_{min} = 1.26 \text{ to } 1.58$

$N_{theo} = 2 N_{min} = 21.8$

$\epsilon = 0.6$

$N_{actual} = \frac{21.8}{0.6}(1.1) = 40 \text{ trays}$

$S_v = 0.1\ kg/h^2$

$u = \frac{1.2 \text{ to } 1.5}{\sqrt{0.1}} = 0.49 \text{ to } 0.6\ m/s$

$D = \left(\frac{4V}{\pi u}\right)^{1/2} = 1.14 \text{ to } 1.48\ m$

$\Delta P_{tower} = N_{actual}(0.007) = 40(0.007) = 0.28\ bar$

f) H-101 fired heater

$Q = 2040\ MJ/h = 751\ kW$

$A_{rad} = \frac{0.5(751)}{37.6} = 99.9\ m^2$

$A_{conv} = \frac{0.5(751)}{12.5} = 300.4\ m^2$

Table 11.2(a) Physical Property Heuristics

	Units	Liquids	Liquids	Gases	Gases	Gases
		Water	Organic Material	Steam	Air	Organic Material
Heat capacity	kJ/kg °C	4.2	1.0–2.5	2.0	1.0	2.0–4.0
Density	kg/m^3	1000	700–1500		1.29@STP	
Latent heat	kJ/kg	1200–2100	200–1000			
Thermal conductivity	W/m °C	0.55–0.70	0.10–0.20	0.025–0.07	0.025–0.05	0.02–0.06
Viscosity	kg/m s	0°C 1.8×10^{-3}	Wide Range	$10–30 \times 10^{-6}$	$20–50 \times 10^{-6}$	$10–30 \times 10^{-6}$
		50°C 5.7×10^{-4}				
		100°C 2.8×10^{-4}				
		200°C 1.4×10^{-4}				
Prandtl no.		1–15	10–1000	1.0	0.7	0.7–0.8

في predict لا يحتاج skip PAR

$Pr = \frac{\text{momentum diffusivity}}{\text{thermal diffusivity}}$

Provides order of magnitude estimates of physical properties used for predication, sanity checks, and preliminary design without detailed property calculations.

Table 11.2(b) Typical Physical Property Variations with Temperature and Pressure

	Liquids	Liquids	Gases	Gases
Property	Temperature	Pressure	Temperature	Pressure
Density	$\rho_l \propto (T_c - T)^{0.3}$	Negligible	$\rho_g = (MW)P/zRT$	$\rho_g = (MW)P/zRT$
Viscosity	$\mu_l = Ae^{B/T}$	Negligible	$\mu_g \propto \frac{T^{1.5}}{(T+1.47T_b)}$	Significant only for $P >$ 10 bar
Vapor pressure	$P^*_c = ae^{b/(T+}$	—	—	—

T is temperature (K), T_c is the critical temperature (K), T_b is the normal boiling point (K), MW is molecular weight, P is pressure, Z is compressibility, R is the gas constant, and P^* is the vapor pressure.

- Density of liquids decreases with temperature
- Density of liquids is almost independent of pressure.
- How does viscosity of liquid change with temperature? Decreases strongly.
- Which property of gases is pressure-sensitive? Density and viscosity at high P

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Table 11.3 Capacities of Process Units in Common Usage[a]

Process Unit	Capacity Unit	Max. Value	Min. Value	Comment
Horizontal vessel	Pressure (bar)	400	Vacuum	L/D typically 2–5, see Table 11.6
	Temper. (°C)	400[b]	−200	
	Height (m)	10	2	
	Diameter (m)	2	0.3	
	L/D	5	2	
Vertical vessel	Pressure (bar)	400	400	L/D typically 2–5, see Table 11.6
	Temper.	400[b]	−200	
		10	2	
	(°C)	2	0.3	
	Height (m)	5	2	
	Diameter (m)			
	L/D			
Towers	Pressure (bar)	400	Vacuum	Normal limits Diameter L/D
	Temper. (°C)	400[b]	−200	0.5 3.0–40[c]
	Height (m)	50	2	1.0 2.5–30[c]
	Diameter (m)	4	0.3	2.0 1.6–23[c]
	L/D	30	2	4.0 1.8–13[c]
Pumps	Power[d] (kW)	250	< 0.1	
Reciprocating	Pressure (bar)	1000	< 0.1	
Rotary and positive Displacement	Power[d] (kW)	150	< 0.1	
Centrifugal	Pressure (bar)	300		
	Power[d] (kW)	250		
	Pressure (bar)	300		
Compressors	Power[d] (kW)	8000	50	
Axial, centrifugal + recipr.	Power[d] (kW)	1000	50	
Rotary				
	(bar)			
	Power[d] (kW)			
	Pressure (bar)			
Compressors	Power[d] (kW)	8000	50	
Axial, centrifugal + recipr.	Power[d] (kW)	1000	50	
Rotary				
Drives for compressors	Power[e] (kW)	15,000	< 1	
Electric	Power[e] (kW)	15,000	100	
Steam turbine	Power[e] (kW)	15,000	10	
Gas turbine	Power[e] (kW)	15,000	10	
Internal combustion eng.	Power[e] (kW)			
Process heaters	Duty (MJ/h)	500,000	10,000	Duties different for reactive heaters/furnaces
Heat exchangers	Area (m^2)	1000	10	For area < 10 m^2 use
	Tube dia. (m)	0.0254	0.019	double pipe exchanger
	Length (m)	6.5	2.5	For 150 < P < 400 bar
	Pressure (bar)	150	Vacuum	need special design
	Temp. (°C)	400[b]	−200	

[a] Most of the limits for equipment sizes shown here correspond to the limits used in the costing program (CAPCOST) introduced in Chapter 7.

[b] Maximum temperature and pressure are related to the materials of construction and may differ from values shown here.

[c] For 20 < L/D < 30 special design may be required. Diameters up to 9 m possible but greater than 4 m must usually be fabricated on site.

[d] Power values refer to fluid/pumping power.

[e] Power values refer to shaft power.

Table 11.4 Effect of Typical Materials of Construction on Product Color, Corrosion,[a] Abrasion, and Catalytic Effects (see also Table 7.9)

Metals		
Material	***Advantages***	***Disadvantages***
Carbon steel	Low cost, readily available, resists abrasion, standard fabrication, resists alkali	Poor resistance to acids and strong alkali, often causes discoloration and contamination
Stainless steel	Resists most acids, reduces discoloration, available with a variety of alloys, abrasion less than mild steel	Not resistant to chlorides, more expensive, fabrication more difficult, alloy materials may have catalytic effects
Monel-Nickel	Little discoloration, contamination, resistant to chlorides	Not resistant to oxidizing environments, expensive
Hasteloy	Improved over Monel-Nickel	More expensive than Monel-Nickel
Other exotic	Improves specific properties	Can be very high cost metals
Nonmetals		
Material	***Advantages***	***Disadvantages***
Glass	Useful in laboratory and batch systems, low diffusion at walls	Fragile, not resistant to high alkali, poor heat transfer, poor abrasion resistance
Plastics	Good at low temperature, large variety to select from with various characteristics, easy to fabricate, seldom discolors, low cost	Poor at high temperature, low strength, not resistant to high-alkali conditions, low heat transfer. Minor catalytic effects possible
Ceramics	Withstands high temperatures, variety of formulations available, modest cost	Poor abrasion properties, high diffusion at walls (in particular hydrogen), low heat transfer, may encourage catalytic reactions

Table 11.5 Heuristics for Drivers and Power Recovery Equipment

1. Efficiency is greater for larger machines. Electric motors are 85%–95%; steam turbines are 42%–78%; gas engines and turbines are 28%–38% efficient (see Figure 8.7).
2. For less than 74.6 kW (100 hp), electric motors are used almost exclusively. They are made for services up to 14,900 kW (20,000 hp).
3. Steam turbines are competitive higher than 76.6 kW (100 hp). They are speed controllable. They are frequently used as spares in case of power failure.
4. Combustion engines and turbines are restricted to mobile and remote locations.
5. Gas expanders for power recovery may be justified at capacities of several hundred horsepower; otherwise any pressure reduction in process is done with throttling valves.
6. The following useful definitions are given:

$$\text{Shaft power} = \frac{\text{theoretical power to pump fluid (liquid or gas)}}{\text{efficiency of pump or compressor, } \varepsilon_{sh}}$$

$$\text{Drive power} = \frac{\text{shaft power}}{\text{efficiency of drive, } \varepsilon_{dr}}$$

$$\text{Overall efficiency} = \varepsilon_{ov} = \varepsilon_{sh}\varepsilon_{dr}$$

ε_{dr} values are given in this table and Figure 8.7.

ε_{sh} values are given in Tables 11.9 and 11.10. Usually ε_{sh} are given on PFD.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment*.

Table 11.6 Heuristics for Process Vessels (Drums)

1. Drums are relatively small vessels that provide surge capacity or separation of entrained phases.
2. Liquid drums are usually horizontal.
3. Gas-liquid phase separators are usually vertical.
4. Optimum ratio of length to diameter = 3, but the range 2.5 to 5 is common.
5. Holdup time is 5 min for half-full reflux drums and gas/liquid separators, 5–10 min for a product feeding another tower.
6. In drums feeding a furnace, 30 min for half-full drum is allowed.
7. Knockout drums placed ahead of compressors should hold no less than 10 times the liquid volume passing per minute.
8. Liquid-liquid separations are designed for settling velocity of 0.085–0.127 cm/s (2–3 in/min).
9. Gas velocity in gas/liquid separators, $u = k\sqrt{\rho_l/\rho_v - 1}$ m/s (ft sec)k = 0.11(0.35) for systems with mesh deentrainer, and k = 0.0305 (0.1) without mesh deentrainer.
10. Entrainment removal of 99% is attained with 10.2–30.5 cm (4–12 in) mesh pad thickness; 15.25 cm (6 in) thickness is popular.
11. For vertical pads, the value of the coefficient in Step 9 is reduced by a factor of 2/3.
12. Good performance can be expected at velocities of 30%–100% of those calculated with the given k; 75% is popular.
13. Disengaging spaces of 15.2–45.7 cm (6–18 in) ahead of the pad and 30.5 cm (12 in) above the pad are suitable.
14. Cyclone separators can be designed for 95% collection at 5 μm particles, but usually only droplets greater than 50 μm need be removed.

See also Chapter 23 for more detailed explanations and guidelines.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.7 Heuristics for Vessels (Pressure and Storage)

Pressure Vessels				
1. Design temperature between −30°C and 345°C is 25°C above maximum operating temperature; higher safety margins are used outside the given temperature range.				
2. The design pressure is 10% or 0.69–1.7 bar (10–25 psi) over the max. operating pressure, whichever is greater. The max. operating pressure, in turn, is taken as 1.7 bar (25 psi) above the normal operation.				
3. Design pressures of vessels operating at 0–0.69 bar (0–10 psig) and 95°C–540°C (200°F–1000°F) are 2.76 barg (40 psig).				
4. For vacuum operation, design pressures are 1 barg (15 psig) and full vacuum.				
5. Minimum wall thickness for rigidity: 6.4 mm (0.25 in) for 1.07 m (42 in) dia., 8.1 mm (0.32 in) for 1.07–1.52 m (42–60 in) dia., and 11.7 mm (0.38 in) for more than 1.52 m (60 in) dia.				
6. Corrosion allowance 8.9 mm (0.35 in) for known corrosive conditions, 3.8 mm (0.15 in) for noncorrosive streams, and 1.5 mm (0.06 in) for steam drums and air receivers.				
7. Allowable working stresses are one-fourth of the ultimate strength of the material.				
8. Maximum allowable stress depends sharply on temperature.				
Temperature: (°F)	−20 to 650	750	850	1000
(°C)	−30 to 345	400	455	540
Low alloy steel SA 203 (psi)	18,759	15,650	9950	2500
(bar)	1290	1070	686	273
Type 302 stainless steel (psi)	18,750	18,750	15,950	6250
(bar)	1290	1290	1100	431

Storage Vessels
1. For less than 3.8 m^3 (1000 gal), use vertical tanks on legs.
2. Between 3.8 and 38 m^3 (1000 and 10,000 gal), use horizontal tanks on concrete supports.
3. Beyond 38 m^3 (10,000 gal) use vertical tanks on concrete pads.
4. Liquids subject to breathing losses may be stored in tanks with floating or expansion roofs for conservation.
5. Freeboard is 15% below 1.9 m^3 (500 gal) and 10% above 1.9 m^3 (500 gal) capacity.
6. Thirty-day capacity often is specified for raw materials and products but depends on connecting transportation equipment schedules.
7. Capacities of storage tanks are at least 1.5 times the size of connecting transportation equipment, for instance, 28.4 m^3 (7500 gal) tanker trucks, 130 m^3 (34,500 gal) rail cars, and virtually unlimited barge and tanker capacities.

Table 11.8 Heuristics for Piping

1. Line velocities (u) and pressure drop (ΔP): (a) for liquid pump discharge: $u = (5 + D/3)$ ft/sec and $\Delta P = 2.0$ psi/100 ft; (b) for liquid pump suction: $u = (1.3 + D/6)$ ft/sec and $\Delta P = 0.4$ psi/100 ft; (c) for steam or gas flow: $u = 20D$ ft/sec and $\Delta P = 0.5$ psi/100 ft, D = diameter of pipe in inches.
2. Gas/steam line velocities = 61 m/s (200 ft/sec), and pressure drop = 0.1 bar/100 m (0.5 psi/100 ft).
3. In preliminary estimates set line pressure drops for an equivalent length of 30 m (100 ft) of pipe between each piece of equipment.
4. Control valves require at least 0.69 bar (10 psi) drop for good control.
5. Globe valves are used for gases, control, and wherever tight shutoff is required. Gate valves for most other services.
6. Screwed fittings are used only on sizes 3.8 cm (1.5 in) or less; otherwise, flanges or welding used.
7. Flanges and fittings are rated for 10, 20, 40, 103, 175 bar (150, 300, 600, 1500, or 2500 psig).
8. Approximate schedule number required = $1000\,P/S$, where P is the internal pressure in psig and S is the allowable working stress [about 690 bar (10,000 psi)] for A120 carbon steel at 260° (500°F). Schedule 40 is most common.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.9 Heuristics for Pumps

1. Power for pumping liquids: kW = (1.67)[Flow(m^3/min)][ΔP(bar)]/ε, [hp = Flow(gpm) ΔP(psi)/1714/ε], ε = Fractional Efficiency = ε_{sh} (see Table 11.5).
2. Net positive suction head (NPSH) of a pump must be in excess of a certain number, depending upon the kind of pumps and the conditions, if damage is to be avoided. *NPSH* = (pressure at the eye of the impeller − vapor pressure)/($\rho\, g$). Common range is 1.2–6.1 m of liquid (4–20 ft).
3. Specific speed N_s = (rpm)(gpm)$^{0.5}$/(head in feet)$^{0.75}$. Pump may be damaged if certain limits on N_s are exceeded, and the efficiency is best in some ranges.
4. Centrifugal pumps: single stage for 0.057–18.9 m^3/min (15–5000 gpm), 152 m (500 ft) maximum head; multistage for 0.076–41.6 m^3/min (20–11,000 gpm), 1675 m (5500 ft) maximum head. Efficiency 45% at 0.378 m^3/min (100 gpm), 70% at 1.89 m^3/min (500 gpm), 80% at 37.8 m^3/min (10,000 gpm).
5. Axial pumps for 0.076–378 m^3/min (20–100,000 gpm), 12 m (40 ft) head, 65%–85% efficiency.
6. Rotary pumps for 0.00378–18.9 m^3/min (1–5000 gpm), 15,200 m (50,000 ft head), 50%–80% efficiency.
7. Reciprocating pumps for 0.0378–37.8 m^3/min (10–10,000 gpm), 300 km (1,000,000 ft) head max. Efficiency 70% at 7.46 kW (10 hp), 85% at 37.3 kW (50 hp), and 90% at 373 kW (500 hp).

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.10 Heuristics for Compressors, Fans, Blowers, and Vacuum Pumps

1. Fans are used to raise the pressure about 3% {12 in (30 cm) water}, blowers to raise less than 2.75 barg (40 psig), and compressors to higher pressures, although the blower range is commonly included in the compressor range.
2. Theoretical reversible adiabatic power = $mz_1RT_1[(\{P_2/P_1\}^a - 1)]/a$ where T_1 is inlet temperature, R = gas constant, z_1 = compressibility, m = molar flow rate, $a = (k-1)/k$, and $k = C_p/C_v$.
3. Values of R: = 8.314 J/mol K = 1.987 Btu/lbmol R = 0.7302 atm ft^3/lbmol R
3. Outlet temperature for reversible adiabatic process $T_2 = T_1 (P_2/P_1)^a$.
4. Exit temperatures should not exceed 167°C–204°C (350°F–400°F); for diatomic gases $C_p/C_v = 1.4$. This corresponds to a compression ratio of about 4.
5. Compression ratio should be about the same in each stage of a multistage unit, ratio = $(P_n/P_1)^{1/n}$, with n stages.
6. Efficiencies of reciprocating compressors: 65% at compression ratios of 1.5, 75% at 2.0, and 80–85% at 3–6.
7. Efficiencies of large centrifugal compressors, 2.83–47.2 m^3/s (6000–100,000 acfm) at suction, are 76–78%.
8. For vacuum pumps use the following:

Reciprocating piston type	Down to 1 Torr
Rotary piston type	Down to 0.001 Torr
Two-lobe rotary type	Down to 0.0001 Torr
Steam jet ejectors	1-stage down to 100 Torr
	3-stage down to 1 Torr
	5-stage down to 0.05 Torr

6. A three-stage ejector needs 100 kg steam/kg air to maintain a pressure of 1 Torr.
7. In-leakage of air to evacuated equipment depends on the absolute pressure, Torr, and the volume of the equipment, V in m^3 (ft^3) according to $W = kV^{2/3}$ kg/h (lb/hr) with k = 0.98 (0.2) when $P > 90$ Torr, k = 0.39 (0.08) between 3 and 20 Torr, and k = 0.12 (0.025) at less than 1 Torr. See Chapter 23 for more details.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.11 Heuristics for Heat Exchangers

1. For conservative estimate set F = 0.9 for shell-and-tube exchangers with no phase changes, $q = UAF\Delta T_{lm}$. When ΔT at exchanger ends differ greatly, then check F, and reconfigure if F is less than 0.85.
2. Standard tubes are 1.9 cm (3/4 in) OD, on a 2.54 cm (1 in) triangle spacing, 4.9 m (16 ft) long.
   A shell 30 cm (1 ft) dia. accommodates 9.3 m^2 (100 ft^2)
   60 cm (2 ft) dia. accommodates 37.2 m^2 (400 ft^2)
   90 cm (3 ft) dia. accommodates 102 m^2 (1100 ft^2)
3. Tube side is for corrosive, fouling, scaling, and high-pressure fluids.
4. Shell side is for viscous and condensing fluids.
5. Pressure drops are 0.1 bar (1.5 psi) for boiling and 0.2–0.62 bar (3–9 psi) for other services.
6. Minimum temperature approach is 10°C (20°F) for fluids and 5°C (10°F) for refrigerants.
7. Cooling water inlet is 30°C (90°F), maximum outlet 45°C (115°F).
8. Heat transfer coefficients for estimating purposes, W/m^2°C (Btu/hr ft^2°F): water to liquid, 850 (150); condensers, 850 (150); liquid to liquid, 280 (50); liquid to gas, 60 (10); gas to gas 30 (5); reboiler 1140 (200). Maximum flux in reboiler 31.5 kW/m^2 (10,000 Btu/hr ft^2).
   When phase changes occur, use a zoned analysis with appropriate coefficient for each zone.
9. Double pipe exchanger is competitive at duties requiring 9.3–18.6 m^2 (100–200 ft^2).
10. Compact (plate and fin) exchangers have 1150 m^2/m^3 (350 ft^2/ft^3), and about 4 times the heat transfer per cut of shell-and-tube units.
11. Plate and frame exchangers are suited to high-sanitation services and are 25%–50% cheaper in stainless steel construction than shell-and-tube units.
12. Air coolers: Tubes are 0.75–1.0 in. OD, total finned surface 15–20 m^2/m^2 (ft^2/ft^2 bare surface), U = 450–570 W/m^2°C (80–100 Btu/hr ft^2 (bare surface) °F). Minimum approach temperature = 22°C (40°F). Fan input power = 1.4–3.6 kW/(MJ/h) [2–5 hp/(1000 Btu/hr)].
13. Fired heaters: Radiant rate, 37.6 kW/m^2 (12,000 Btu/hr ft^2); convection rate, 12.5 kW/m^2 (4000 Btu/hr ft^2); cold oil tube velocity = 1.8 m/s (6 ft/sec); approximately equal transfer in the two sections; thermal efficiency 70%–90% based on lower heating value; flue gas temperature 140°C–195°C (250°F–350°F) above feed inlet; stack gas temperature 345°C–510°C (650°F–950°F).

Reciprocating → high P high eff

Centrifugal → large flow rates

surity check
Ratio = 10

multi stage

To limit outlet T and reduce Power Consumption

Small ΔP → fan
medium ΔP → blower
high ΔP → compressor

Table 11.12 Heuristics for Thermal Insulation

1. Up to 345°C (650°F), 85% magnesia is used.
2. Up to 870°C–1040°C (1600°F–1900°F), a mixture of asbestos and diatomaceous earth is used.
3. Ceramic (refractory) linings at higher temperature.
4. Cryogenic equipment −130°C (−200°F) employs insulation with fine pores of trapped air, e.g., Perlite.
5. Optimal thickness varies with temperature: 1.27 cm (0.5 in) at 95°C (200°F), 2.54 cm (1.0 in) at 200°C (400°F), 3.2 cm (1.25 in) at 315°C (600°F).
6. Under windy conditions 12.1 km/h (7.5 miles/hr), 10%–20% greater thickness of insulation is justified.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.13 Heuristics for Towers (Distillation and Gas Absorption)

1. Distillation is usually the most economical method for separating liquids, superior to extraction, absorption crystallization, or others.
2. For ideal mixtures, relative volatility is the ratio of vapor pressures $\alpha_{12} = P_1^*/P_2^*$.
3. Tower operating pressure is most often determined by the temperature of the condensing media, 38°C–50°C (100°F–120°F) if cooling water is used, or by the maximum allowable reboiler temperature to avoid chemical decomposition/degradation.
4. Sequencing of columns for separating multicomponent mixtures:[a]
   1. Perform the easiest separation first, that is, the one least demanding of trays and reflux, and leave the most difficult to the last.
   2. When neither relative volatility nor feed composition varies widely, remove components one by one as overhead products.
   3. When the adjacent ordered components in the feed vary widely in relative volatility, sequence the splits in order of decreasing volatility.
   4. When the concentrations in the feed vary widely but the relative volatilities do not, remove the components in order of decreasing concentration.
5. Economical optimum reflux ratio is in the range of 1.2 to 1.5 times the minimum reflux ratio, R_{min}.
6. The economically optimum number of theoretical trays is near twice the minimum value N_{min}.
7. The minimum number of trays is found with the Fenske-Underwood equation:
   $N_{min} = \ln\{[x/(1-x)]_{ovhd}/[x/(1-x)]_{btms}\}/\ln \alpha$
8. Minimum reflux for binary or pseudobinary mixtures is given by the following when separation is essentially complete ($x_D = 1$) and D/F is the ratio of overhead product to feed rate: $R_{min}D/F = 1/(\alpha - 1)$, when feed is at the bubble point
   $(R_{min} + 1)\ D/F = \alpha/(\alpha - 1)$, when feed is at the dew point
9. A safety factor of 10% of the number of trays calculated by the best means is advisable.
10. Reflux pumps are made at least 10% oversize.
11. The optimum value of the Kremser absorption factor $A = (L/mV)$ is in the range of 1.25 to 2.0.
12. Reflux drums usually are horizontal, with a liquid holdup of 5 min half-full. A takeoff pot for a second liquid phase, such as water in hydrocarbon systems, is sized for a linear velocity of that phase of 1.3 m/s (0.5 ft/sec), minimum diameter is 0.4 m (16 in).
13. For towers about 0.9 m (3 ft) dia., add 1.2 m (4 ft) at the top for vapor disengagement, and 1.8 m (6 ft) at the bottom for liquid level and reboiler return.
14. Limit the tower height to about 53 m (175 ft) max. because of wind load and foundation considerations. An additional criterion is that L/D be less than 30 ($20 < L/D < 30$ often will require special design).

[a] Additional information on sequencing is given in Table 12.2.

Table 11.14 Heuristics for Tray Towers (Distillation and Gas Absorption)

1. For reasons of accessibility, tray spacings are made 0.5–0.6 m (20–24 in).
2. Peak efficiency of trays is at values of the vapor factor $F_s = u\rho^{0.5}$ in the range of 1.2–1.5 m/s $\{kg/m^3\}^{0.5}$ [1–1.2 ft/s $\{lb/ft^3\}^{0.5}$]. This range of F_s establishes the diameter of the tower. Roughly, linear velocities are 0.6 m/s (2 ft/sec) at moderate pressures, and 1.8 m/s (6 ft/sec) in vacuum.
3. Pressure drop per tray is on the order of 7.6 cm (3 in) of water or 0.007 bar (0.1 psi).
4. Tray efficiencies for distillation of light hydrocarbons and aqueous solutions are 60%–90%; for gas absorption and stripping, 10%–20%.
5. Sieve trays have holes 0.6–0.7 cm (0.25–0.5 in) dia., area being 10% of the active cross section.
6. Valve trays have holes 3.8 cm (1.5 in) dia. each provided with a liftable cap, 130–150 caps/m^2 (12–14 caps/ft^2) of active cross section. Valve trays are usually cheaper than sieve trays.
7. Bubblecap trays are used only when a liquid level must be maintained at low turndown ratio; they can be designed for lower pressure drop than either sieve or valve trays.
8. Weir heights are 5 cm (2 in), weir lengths are about 75% of tray diameter, liquid rate—a maximum of 1.2 m^3/min m of weir (8 gpm/in of weir); multipass arrangements are used at higher liquid rates.

Table 11.15 Heuristics for Packed Towers (Distillation and Gas Absorption)

1. Structured and random packings are suitable for packed towers less than 0.9 m (3 ft) when low pressure drop is required.
2. Replacing trays with packing allows greater throughput and separation in existing tower shells.
3. For gas rates of 14.2 m^3/min (500 ft^3/min), use 2.5 cm (1 in) packing; for 56.6 m^3/min (2000 ft^3/min) or more, use 5 cm (2 in) packing.
4. Ratio of tower diameter to packing diameter should be >15:1.
5. Because of deformability, plastic packing is limited to 3–4 m (10–15 ft) and metal to 6.0–7.6 m (20–25 ft) unsupported depth.
6. Liquid distributors are required every 5–10 tower diameters with pall rings, and at least every 6.5 m (20 ft) for other types of dumped packing.
7. Number of liquid distributors should be >32–55/m^2 (3–5/ft^2) in towers greater than 0.9 m (3 ft) diameter, and more numerous in smaller columns.
8. Packed towers should operate near 70% of flooding (evaluated from Sherwood and Lobo correlation).
9. Height equivalent to theoretical stage (HETS) for vapor-liquid contacting is 0.4–0.56 m (1.3–1.8 ft) for 2.5 cm (1 in) pall rings, and 0.76–0.9 m. (2.5–3.0 ft) for 5 cm (2 in) pall rings.
10. Generalized pressure drops

	Design Pressure Drops (cm of H_2O/m of packing)	Design Pressure Drops (inches of H_2O/ft of packing)
Absorbers and regenerators (nonfoaming systems)	2.1–3.3	0.25–0.40
Absorbers and regenerators	0.8–2.1	0.10–0.25
Atmospheric/pressure stills and fractionators	3.3–6.7	0.40–0.80
Vacuum stills and fractionators	0.8–3.3	0.10–0.40
Maximum value	8.33	1.0

Table 11.16 Heuristics for Liquid-Liquid Extraction

1. The dispersed phase should be the one with the higher volumetric flowrate except in equipment subject to back-mixing, where it should be the one with the smaller volumetric rate. It should be the phase that wets material of construction less well. Because the holdup of continuous phase is greater, that phase should be made up of the less expensive or less hazardous material.
2. There are no known commercial applications of reflux to extraction processes, although the theory is favorable.
3. Mixer-settler arrangements are limited to at most five stages. Mixing is accomplished with rotating impellers or circulation pumps. Settlers are designed on the assumption that droplet sizes are about 150 μm dia. In open vessels, residence times of 30–60 min or superficial velocities of 0.15–0.46 m/min (0.5–1.5 ft/min) are provided in settlers. Extraction stage efficiencies commonly are taken as 80%.
4. Spray towers as tall as 6–12 m (20–40 ft) cannot be depended on to function as more than a single stage.
5. Packed towers are employed when 5–10 stages suffice. Pall rings 2.5–3.8 cm (1–1.5 in) size are best. Dispersed phase loadings should not exceed 10.2 m^3/min m^2 (25 gal/min ft^2). HETS of 1.5–3.0 m (5–10 ft) may be realized. The dispersed phase must be redistributed every 1.5–2.1 m (5–7 ft). Packed towers are not satisfactory when the surface tension is more than 10 dyne/cm.
6. Sieve tray towers have holes of only 3–8 mm dia. Velocities through the holes are kept less than 0.24 m/s (0.8 ft/sec) to avoid formation of small drops. Redispersion of either phase at each tray can be designed for. Tray spacings are 15.2 to 60 cm (6 to 24 in). Tray efficiencies are in the range of 20%–30%.
7. Pulsed packed and sieve tray towers may operate at frequencies of 90 cycles/min and amplitudes of 6–25 mm. In large-diameter towers, HETS of about 1 m have been observed. Surface tensions as high as 30–40 dyne/cm have no adverse effect.
8. Reciprocating tray towers can have holes 1.5 cm (9/16 in) dia., 50%–60% open area, stroke length 1.9 cm (0.75 in), 100–150 strokes/min, plate spacing normally 5 cm (2 in) but in the range of 2.5–15 cm (1–6 in). In a 76 cm (30 in) diameter tower, HETS is 50–65 cm (20–25 in) and throughput is 13.7 m^3/min m^2 (2000 gal/hr ft^2). Power requirements are much less than that of pulsed towers.
9. Rotating disk contactors or other rotary agitated towers realize HETS in the range of 0.1–0.5 m (0.33–1.64 ft). The especially efficient Kuhni with perforated disks of 40% free cross section has HETS of 0.2 m (0.66 ft) and a capacity of 50 m^3/m^2 h (164 ft^3/ft^2 hr).

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment,*

Table 11.17 Heuristics for Reactors

1. The rate of reaction in every instance must be established in the laboratory, and the residence time or space velocity and product distribution eventually must be found from a pilot plant.
2. Dimensions of catalyst particles are 0.1 mm (0.004 in) in fluidized beds, 1 mm in slurry beds, and 2–5 mm (0.078–0.197 in) in fixed beds.
3. The optimum proportions of stirred tank reactors are with liquid level equal to the tank diameter, but at high pressures slimmer proportions are economical.
4. Power input to a homogeneous reaction stirred tank is 0.1–0.3 kW/m^3 (0.5–1.5 hp/1000 gal), but three times this amount when heat is to be transferred.
5. Ideal CSTR (continuous stirred tank reactor) behavior is approached when the mean residence time is 5 to 10 times the length needed to achieve homogeneity, which is accomplished with 500–2000 revolutions of a properly designed stirrer.
6. Batch reactions are conducted in stirred tanks for small daily production rates or when the reaction times are long or when some condition such as feed rate or temperature must be programmed in some way.
7. Relatively slow reactions of liquids and slurries are conducted in continuous stirred tanks. A battery of four or five in series is most economical.
8. Tubular flow reactors are suited to high production rates at short residence times (seconds or minutes) and when substantial heat transfer is needed. Embedded tubes or shell-and-tube construction then is used.
9. In granular catalyst packed reactors, the residence time distribution is often no better than that of a five-stage CSTR battery.
10. For conversion less than about 95% of equilibrium, the performance of a five-stage CSTR battery approaches plug flow.
11. The effect of temperature on chemical reaction rate is to double the rate every 10°C.
12. The rate of reaction in a heterogeneous system is more often controlled by the rate of heat or mass transfer than by the chemical reaction kinetics.
13. The value of a catalyst may be to improve selectivity more than to improve the overall reaction rate.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.18 Heuristics for Refrigeration and Utility Specifications

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Elsevier, Boston, 2012.

Table 11.18 Heuristics for Refrigeration and Utility Specifications

1. A ton of refrigeration is the removal of 12,700 kJ/h (12,000 Btu/hr) of heat.
2. At various temperature levels: −18°C to −10°C (0°F to 50°F), chilled brine and glycol solutions; −45°C to −10°C (−50°F to −40°F), ammonia, freon, butane; −100°C to −45°C (−150°F to −50°F) ethane or propane.
3. Compression refrigeration with 38°C (100°F) condenser requires kW/tonne (hp/ton) at various temperature levels; 0.93 (1.24) at −7°C (20°F); 1.31 (1.75) at −18°C (0°F); 2.3 (3.1) at −40°C (−40°F); 3.9 (5.2) at −62°C (−80°F).
4. At less than −62°C (−80°F), cascades of two or three refrigerants are used.
5. In single-stage compression, the compression ratio is limited to 4.
6. In multistage compression, economy is improved with interstage flashing and recycling, so-called economizer operation.
7. Absorption refrigeration: ammonia to −34°C (−30°F); lithium bromide to 7°C (+45°F) is economical when waste steam is available at 0.9 barg (12 psig).
8. Steam: 1–2 barg (15–30 psig), 121°C–135°C (250°F–275°F); 10 barg (150 psig), 186°C (366°F); 27.6 barg (400 psig), 231°C (448°F); 41.3 barg (600 psig), 252°C (488°F) or with 55°C–85°C (100°F–150°F) superheat.
9. Cooling water: For design of cooling tower use supply at 27°C–32°C (80°F–90°F) from cooling tower, return at 45°C–52°C (115°F–125°F); return seawater at 43°C (110°F); return tempered water or steam condensate above 52°C (125°F).
10. Cooling air supply at 29°C–35°C (85°F–95°F); temperature approach to process, 22°C (40°F).
11. Compressed air 3.1 (45), 10.3 (150), 20.6 (300), or 30.9 barg (450 psi) levels.
12. Instrument air at 3.1 barg (45 psig), −18°C (0°F) dew point.
13. Fuels: gas of 37,200 kJ/m^3 (1000 Btu/SCF) at 0.35–0.69 barg (5–10 psig), or up to 1.73 barg (25 psig) for some types of burners; liquid at 39.8 GJ/m^3 (6 million Btu/bbl).
14. Heat transfer fluids: petroleum oils less than 315°C (600°F), Dowtherms less than 400°C (750°F), fused salts less than 600°C (1100°F), direct fire or electricity above 450°F.
15. Electricity: 0.75–74.7 kW. (1–100 hp), 220–550 V; 149–1864 kW (200–2500 hp), 2300–4000 V.

Source: Adapted from Couper, J. R., et al., *Chemical Process Equipment, Selection and Design*, 3rd ed., Boston: Elsevier, 2012.

WHAT YOU SHOULD HAVE LEARNED

Problems

1. For the ethylbenzene process shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics: P-301, V-302, T-302. Comment on any significant differences that you find.

2. For the styrene process shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics: E-401, C-401, T-402. Comment on any significant differences that you find.

3. For the drying oil shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics: V-501, P-501, H-501. Comment on any significant differences that you find.

Chapter 11

11.1 For the ethylbenzene process shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics, P-301, V-302, T-302. Comment on any significant differences that you find.

P-301 – From Table B.2.3

P-301 A/B, Carbon steel – positive displacement, Actual power = 15 kW, Efficiency 75%

From Table 11.9, Heuristic 1

Power for pumping liquids: kW = (1.67)[Flow(m3/min)][ΔP(bar)]/ε

ε = Fractional Efficiency = ε_{sh} (see Table 11.5)

density of benzene at 58.5°C = 875 kg/m^3

mass flow of benzene through pump = 17952.2 kg/h (Stream 3 – Table B.3.1)

Flowrate = 17952.2/875 = 20.5 m^3/h = 0.342 m3/min

ΔP = 2000 – 110 = 1890 kPa = 18.9 bar

$P_{theoretical}$ = (1.67)(0.342)(18.9) = 10.8 kW

From Table 11.9 - Heuristic 7

Efficiency 70% at 7.46 kW (10 hp), 85% at 37.3 kW (50 hp)

$\varepsilon_{sh} \sim 75\%$

P_{actual} = **10.8/0.75 = 14.4 kW**

Good agreement with heuristics

V-302 - From Table B.2.3

V-302, Carbon steel with SS demister, vertical, L/D = 3, V = 10 m^3, Maximum operating pressure = 250 kPa

From Table 11.6 – Heuristic 4

Optimum length/diameter = 3, but the range 2.5 to 5 is common.

Use L/D = 3

From Table 11.6 – Heuristic 5

Holdup time is 5 min for half-full reflux drums and gas/liquid separators, 5–10 min for a product feeding another tower.

Since L/V separator feeds a tower use a hold up of liquid equal to 10 min for half-full drum

Properties of Stream 15	Properties of Stream 16
Vapor	Liquid
Flowrate = 1038 kg/h	Flowrate = 24345.9 kg/h
Density = 2.169 kg/m^3	Density = 821.3 kg/m^3
Vol flow = 478.6 m^3/h	Vol flow = 29.64 m^3/h

Volume of drum, $V = \dfrac{(29.64)(10)}{(60)(0.5)} = 10\,\text{m}^3$

Good agreement with heuristics

T-302 – From TableB.2.3

T-302, carbon steel, 76 SS sieve trays plus reboiler and total condenser, 45% efficient trays, feed on tray 56, additional feeds ports on 50 and 62, reflux ratio = 0.6608, 15 in tray spacing, column height 28.96 m, diameter = 1.5 m, maximum pressure rating of 300 kPa

From Tables 11.13 and 11.14– Heuristics

Table 11.13
Rule 5: Optimum reflux in the range of 1.2 – 1.5 R_{min}
Rule 6: Optimum number of stages approximately $2N_{min}$
Rule 7: $N_{min} = \ln\{ [x/(1-x)]_{ovhd}/[x/(1-x)]_{bot}\}/\ln \alpha$
Rule 8: $R_{min} = \{F/D\}/(\alpha - 1)$
Rule 9: Use a safety factor of 10% on number of trays

Table 11.14 → Rule 2: $Fs = u\rho_v^{0.5} = 1.2 \rightarrow 1.5$ m/s(kg/m3)$^{0.5}$ and Rule 4: $\varepsilon_{tray} = 60 - 90$ %

For our case we have

Stream	18	19	20
Temp °C	145.4	139.0	191.1
Pres kPa	120.0	110.0	140.0
Vapor mole fraction	0.0	0.0	0.0
Total kmol/h	101.1	89.9	11.3
Total kg/h	11024.5	9538.6	1485.9
Flowrates in kmol/h			
Ethylene	0.00	0.00	0.00
Ethane	0.00	0.00	0.00
Propylene	0.00	0.00	0.00
Benzene	0.17	0.17	0.00
Toluene	0.00	0.00	0.00
Ethylbenzene	90.63	89.72	0.91
1,4-DiEthBenzene	10.35	0.0001	10.35

Key components are ethylbenzene and 1,4-DiEthylbenzene. The formulae for N_{min}, etc. should be based on [redacted].

Propylene			0.00
Benzene	0.17	0.17	0.00
Toluene	0.00	0.00	0.00
Ethylbenzene	90.63	89.72	0.91
1,4-DiEthBenzene	10.35	0.0001	10.35

Key components are ethylbenzene and 1,4-DiEthylbenzene. The formulae for N_{min}, etc. should be based on these key components, i.e., use a benzene free basis.

$x_{ovhd} = 89.72/89.7201 = 0.999999$, $x_{bot} = 0.91/11.3 = 0.08053$, $\alpha_{ovhd} = 3.83$, $\alpha_{bot} = 3.19$
$\alpha_{geom\ ave} = (\alpha_{ovhd}\alpha_{bot})^{0.5} = 3.50$
$N_{min} = \ln\{ [0.999999/(1 - 0.999999)]/[0.08053/(1 - 0.08053)]\} / \ln (3.50) = 12.9$
$R_{min} = \{101.1/89.9\}/(3.5 - 1) = 0.453$
Range of $R = (1.2 \rightarrow 1.5)R_{min} = 0.544 \rightarrow 0.680$

$\boxed{R_{T\text{-}302} = 0.6608}$ Within range

$N_{theoretical} \approx (2)(12.9) = 25.8$
$\varepsilon_{tray} = 0.60 - 0.90$

$\boxed{\varepsilon_{T\text{-}302} = 0.45}$ Lower than typical range

$N_{actual} \approx (25.8/0.45)(1.1) = 63$ trays

$\boxed{N_{T\text{-}302} = 76}$ High by 13 trays

$\rho_v = 3.546 - 3.879 \rightarrow$ use 3.71 kg/m3
$u = (1.2 \rightarrow 1.5)/3.71^{0.5} = 0.62 \rightarrow 0.78$ m/s
Vapor flow rate (stream 19) = 9539 kg/h
Vol. flow rate, $v = 0.714$ m^3/s
$D_{tower} = [4v/\pi u]^{0.5} = [(4)(.714)/(3.142)/(0.62 \rightarrow 0.78)]^{0.5} = 1.08 \rightarrow 1.21$ m

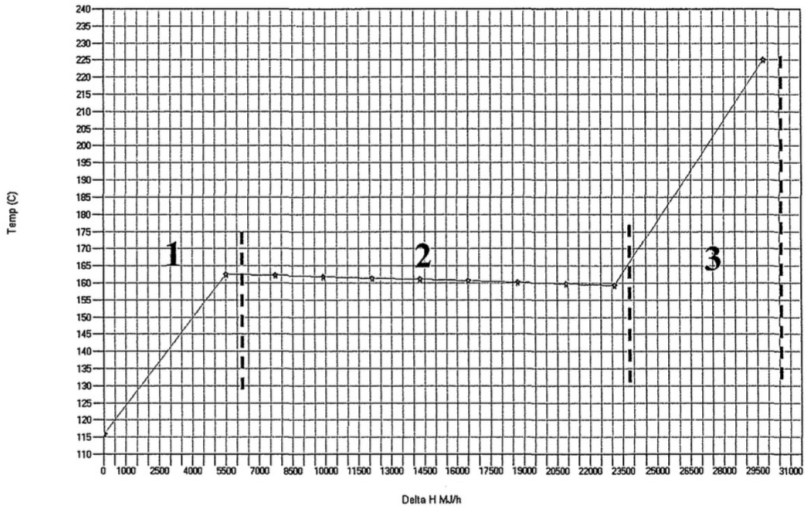
$\boxed{D_{T\text{-}302} = 1.5\ \text{m}}$ Higher than above range for D

Overall, for T-302, the agreement with the heuristics is fair. This is probably due, in part, to the use of a fairly low reflux ratio that is at the bottom of the typical range given in the heuristics. This tends to increase the number of trays. In addition, the tray spacing used in T-302 is only 12 or 15" which is significantly lower than the standard spacing of 24". This has the effect of reducing the tray efficiency and requires a larger diameter column to stay away from flooding. These differences probably account for the discrepancies in the above results.

11.2 For the styrene process shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics, E-401, C-401, T-402. Comment on any significant differences that you find.

E-401 – From Table B.3.3
carbon steel, A = 541 m^2, boiling in shell, condensing in tubes, 1 shell – 2 tube passes, Q = 29,695 MJ/h

From Table B.3.1 and Figure B.3.1, it can be seen that E-401 heats Stream 2 from 116°C to 225°C using high-pressure steam at 254°C. The T-Q diagram from Chemcad is given below:



This is a 3 zone HX.

Q_1 = 5,500 MJ/h
Q_2 = 23,000 – 5,500 = 17,500 MJ/h
Q_3 = 29,700 – 23,000 = 6,700 MJ/h

From Table 11.11 – Heuristic 8
Heat transfer coefficients for estimating purposes, W/m^2°C: water to liquid, 850; condensers, 850; liquid to liquid, 280; liquid to gas, 60; gas to gas 30; reboiler 1140. Maximum flux in reboiler 31.5 kW/m^2.
So choose U_1 = 500 (liq – cond steam). U_2 = 850 (boiling liq – condensing steam, U_3 = 60 (condensing steam – gas)

$$\Delta T_{lm-1} = \frac{[(254-115)-(254-163)]}{\ln\frac{(254-115)}{(254-163)}} = 113.3^{\circ}\text{C}$$

$$\Delta T_{lm-2} = (254-163) = 91^{\circ}\text{C}$$

$$\Delta T_{lm-3} = \frac{[(254-163)-(254-225)]}{\ln\frac{(254-163)}{(254-225)}} = 54.2^{\circ}\text{C}$$

$$\Delta T_{lm-1} = \frac{[(254-115)-(254-163)]}{\ln\frac{(254-115)}{(254-163)}} = 113.3^{o}\text{C}$$

$$\Delta T_{lm-2} = (254-163) = 91^{o}\text{C}$$

$$\Delta T_{lm-3} = \frac{[(254-163)-(254-225)]}{\ln\frac{(254-163)}{(254-225)}} = 54.2^{o}\text{C}$$

Using a value of F = 1 – since all zones have a phase change, we get

$$A_1 = \frac{Q_1}{U_1 \Delta T_{lm-1}} = \frac{(5{,}500\times10^6)}{(3600)(500)(113.3)} = 27.0\text{m}^2$$

$$A_2 = \frac{Q_2}{U_2 \Delta T_{lm-2}} = \frac{(17{,}500\times10^6)}{(3600)(850)(91)} = 62.8\text{m}^2$$

$$A_3 = \frac{Q_3}{U_3 \Delta T_{lm-3}} = \frac{(6{,}700\times10^6)}{(3600)(60)(54.2)} = 572.3\text{m}^2$$

$$A_{total} = \sum_{i=1}^{3} A_i = 662.1\text{m}^2$$

$$\boxed{A_{E\text{-}401} = 541\ \text{m}^2}$$

Agreement is within 20% of heuristic - OK

C-401– From Table B.3.3
carbon steel, W = 364.2 kW, 80% adiabatic efficiency

Feed stream = Stream 16
T = 65°C, P = 0.75 bar, m = 2682 kg/h, mw = 12.4 kg/kmol
P_{out} = 2.4 bar

From Table 11.10 – Heuristic 2
Theoretical reversible adiabatic power = $mz_1RT_1[(\{P_2/P_1\}^a - 1)]/a$
where T_1 is inlet temperature, R = Gas Constant, z_1 = compressibility, m = molar flow rate, $a = (k\text{-}1)/k$ and $k = C_p/C_v$, R = 8.314 J/mol K

assume k = 1.4, a = 0.2857

$$P_{rev-adiab} = \frac{2682\times10^3}{(12.4)(3600)}\frac{(1)(8.314)(273.2+65)}{0.2857}\left[\left(\frac{2.4}{0.75}\right)^{0.2857} - 1\right] = 233085\text{W} = 233\text{kW}$$

From Table 11.10 – Heuristic 8
Efficiencies of large centrifugal compressors is about 76 – 78%

P_{actual} = 233/0.77 = 303 kW

$\eta_{C\text{-}401} = 0.80$

$$P_{rev-adiab} = \frac{2682\times10^3}{(12.4)(3600)} \frac{(1)(8.314)(273.2+65)}{0.2857}\left[\left(\frac{2.4}{0.75}\right)^{0.2857} - 1\right] = 233085\text{W} = 233\text{kW}$$

From Table 11.10 – Heuristic 8
Efficiencies of large centrifugal compressors is about 76 – 78%

P_{actual} = 233/0.77 = 303 kW

$\varepsilon_{C\text{-}401}$ = **0.80**
$P_{C\text{-}401}$ = **364 kW**

Compressor appears to be somewhat oversized (~20%), although the estimate is in the ballpark. This may be to allow for future expansion or could be an error in the design calculations.

T-402– From Table B.3.3
carbon steel, total condenser (E-409), feed at location equivalent to tray 36, reflux ratio = 25.8, structured packing, C_f = 1, diameter = 4.1 m, *HETP* = 0.3 m, height = 34.5 m

From Tables 11.13 and 11.14– Heuristics

Table 11.13
Rule 5: Optimum reflux in the range of 1.2 − 1.5 R_{min}
Rule 6: Optimum number of stages approximately $2N_{min}$
Rule 7: $N_{min} = \ln\{[x/(1-x)]_{ovhd}/[x/(1-x)]_{bot}\}/\ln\alpha$
Rule 8: $R_{min} = \{F/D\}/(\alpha - 1)$
Rule 9: Use a safety factor of 10% on number of trays

Table 11.14
Rule 2: $Fs = u\rho_v^{0.5} = 1.2 \rightarrow 1.5$ m/s(kg/m3)$^{0.5}$
Rule 4: ε_{tray} = 60 – 90 %

From Table 11.15 – Heuristic 9
Height equivalent to theoretical stage (HETS) for vapor-liquid contacting is 0.4-0.56 m for 2.5 cm (1 in) pall rings and 0.76–0.9 m. for 5 cm (2 in) pall rings.

Structured packing is generally more efficient than loose packing so an HETP (HETS) = 0.3m is reasonable.
Assuming a headspace of 1.5 m above the packing and a liquid level of 3m at the bottom

Number of equivalent theoretical plates = (34.5 – 4.5)/0.3 = 100 plates

Using data from Table B.3.3 for the feed and product streams for T-402, we get the following

following

Stream No.	22	23	24
Temperature (°C)	119.5	105	124.5
Pressure (kPa)	60	210	60
Vapor Mole Fraction	0	0	0
Total Flow (kg/h)	47905	35473	12432
Total Flow (kmol/h)	453.9	334.2	119.7
Component Flows			
Water			
Ethylbenzene	333.0	332.66	0.34
Styrene	120.53	1.20	119.3
Hydrogen			
Benzene			
Toluene	0.33	0.33	
Ethylene			
Methane			

Using a toluene free basis, we have

$x_{ovhd} = 332.66/(332.66+1.20) = 0.9964$
$x_{bot} = (0.34)/(119.3 + 0.34) = 0.002842$
From the Chemcad output we get the following values for relative volatility

$\alpha_{top} = 1.19$
$\alpha_{bot} = 1.22$
$\alpha_{ave} = [(1.19)(1.22)]^{1/2} = 1.205$
$N_{min} = \ln\{ [x/(1-x)]_{ovhd}/[x/(1-x)]_{bot}\}/\ln \alpha$

$$N_{min} = \frac{\ln \frac{x_{ovhd}(1-x_{bot})}{(1-x_{ovhd})x_{bot}}}{\ln \alpha_{ave}} = \frac{\ln \frac{(0.9964)(1-0.002842)}{(1-0.9964)(0.002842)}}{\ln(1.205)} = 61.7$$

$R_{min} = \{F/D\}/(\alpha - 1)$

$$R_{min} = \frac{(453.9)}{(334.2)(1.205-1)} = 6.63$$

Using heuristics we get
$R = (1.2 - 1.5)R_{min} = 7.95 - 9.94$
$N = 2N_{min} = 123$

$$\boxed{\begin{array}{l} R_{T\text{-}402} = 25.8 \\ N_{T\text{-}402} = 100 \end{array}}$$

The reflux ratio for this column is much higher than the value from the heuristic while the number of trays is lower by ~20%. This column appears to be running inefficiently and should be investigated since the reboiler and condenser duties are probably very high and wasteful in utilities. One possible explanation is that the higher value of *R* is chosen to keep the height of the tower down but this seems to be overkill. Tower needs investigating!

11.3 For the drying oil shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristic V-501, P-501, H-501. Comment on any significant differences that you find.

11.3 For the drying oil shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristic V-501, P-501, H-501. Comment on any significant differences that you find.

P-501 A/B – From Table B.4.3
Centrifugal, Carbon steel, Power = 0.9 kW (actual), 80% efficient, $NPSH_R$ at design flow = 14 ft of liquid

From Table 11.9

1. Power for pumping liquids: kW = (1.67)[Flow(m^3/min)][ΔP(bar)]/ε, ε = Fractional Efficiency = ε_{sh} (see Table 11.5)
2. Net positive suction head (NPSH) of a pump must be in excess of a certain number, depending upon the kind of pumps and the conditions, if damage is to be avoided. $NPSH$ = (pressure at the eye of the impeller -vapor pressure)/(ρg). Common range is 1.2–6.1 m of liquid
4. Centrifugal pumps: Single stage for 0.057–18.9 m^3/min , 152 m maximum head; multistage for 0.076–41.6 m^3/min , 1675 m maximum head. Efficiency 45% at 0.378 m^3/min, 70% at 1.89 m^3/min , 80% at 37.8 m^3/min.

From Table 11.8
4. Control valves require at least 0.69 bar (10 psi) drop for good control.

Inlet stream is Stream 2 from Figure B.4.1 and Table B.4.1

Properties of Stream 2
Liquid
Flowrate = 10,703 kg/h
Density = 793.5 kg/m^3
Vol flow = 13.49 m^3/h = 0.2248 m^3/min
P_2 = 105 kPa = 1.05 bar
P_3 = 230 kPa = 2.30 bar (Stream 3)
Note: $\Delta P_{pump} = P_3 - P_2 + \Delta P_{CV}$

$$P = \frac{1.67V\Delta P}{\varepsilon} = \frac{(1.67)(13.49/60)(2.30-1.05+0.69)}{\varepsilon} = \frac{0.728}{\varepsilon}\text{kW}$$

From Table 11.9 – heuristic 4, the efficiency is ~ 45%. This is much lower than the actual efficiency of 80%. Using ε = 80% we get:

$$P = \frac{0.728}{0.8} = 0.91\text{kW}$$

Excellent agreement with actual power, but the efficiency looks to be too high for such a small pump. This may be a mistake or a high efficiency pump has been specified for this service.

H-501 – From Table B.4.3
total heat duty required = 13219 MJ/h = 3672 kW, design capacity = 4000 kW, Carbon steel tubes, 85% thermal efficiency

From Table 11.11 – Heuristic 13
Fired heaters: radiant rate, 37.6 kW/m^2; convection rate, 12.5 kW/m^2; cold oil tube velocity = 1.8 m/s approximately equal transfer in the two sections; thermal efficiency 70–90% based on lower heating value; flue gas temperature 140–195°C above feed inlet; stack gas temperature 345–510°C

Only thing to check here is the thermal efficiency which lies in the acceptable range, namely 70%<85%<90% - **good agreement with heuristics**.

V-501 - From Table B.4.3
V-501, Horizontal, Carbon steel, L/D = 3, V = 2.3 m^3

From Table 11.6 – Heuristic 4

Optimum length/diameter = 3, but the range 2.5 to 5 is common.

Use $L/D = 3$

From Table 11.6 – Heuristic 5
Holdup time is 5 min for half-full reflux drums

Properties of Stream 2
Liquid
Flowrate = 10,703 kg/h
Density = 793.5 kg/m^3
Vol flow = 13.49 m^3/h

Volume of drum, $V = \dfrac{(13.49)(5)}{(60)(0.5)} = 2.25\,\text{m}^3$

Good agreement with heuristics

Ch 15:

Pinch technology : is a systematic method used to minimize energy consumption by maximizing internal heat or mass recovery.

Pinch point : Is the location in a system where the driving force for heat or mass transfer is at is minimum.

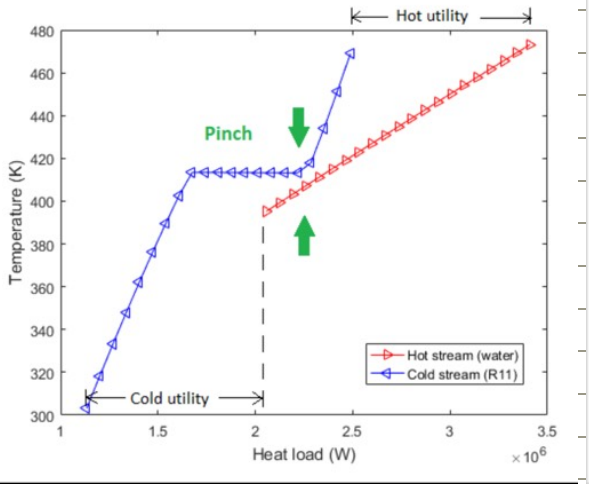
As the driving force decreases, the required equipment size and cost increase.

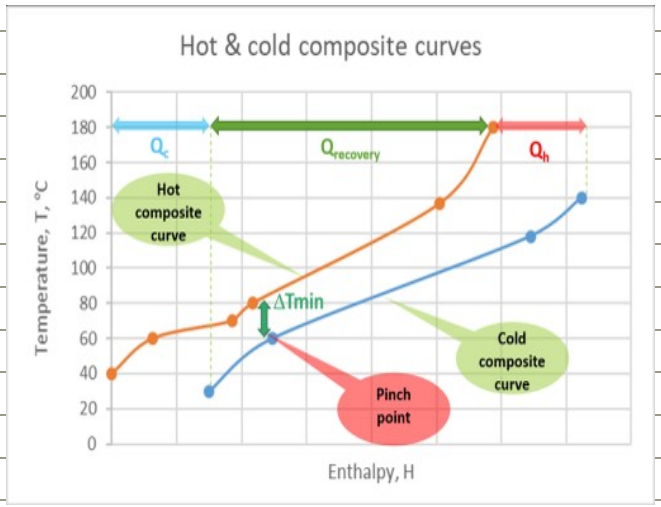
The pinch occurs when a system approaches thermodynamic limits imposed by physical laws.

In heat exchangers, a very small temperature approach requires a very large heat transfer area.

Heat integration means using waste heat from hot streams to heat cold streams within the process.

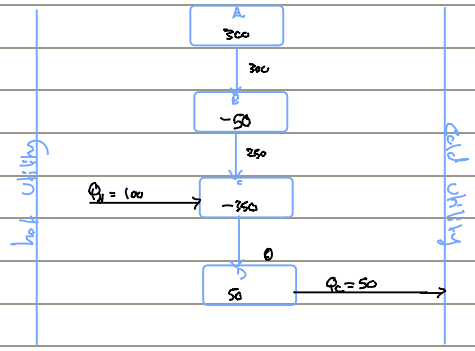
Heat exchanger network (HEN) that transfer between hot and cold stream.





Ex. 15.2.

Stream	Condition	Flowrate, $\dot{m}$ (kg/s)	C_p (kJ/kg°C)	$\dot{m}C_p$ (kW/°C)	T_{in} (°C)	T_{out} (°C)	$Q_{available}$ (kW)
1	Hot	10.00	0.8	8.0	300	150	1200
2	Hot	2.50	0.8	2.0	150	50	200
3	Hot	3.00	1.0	3.0	200	50	450
4	Cold	6.25	0.8	5.0	190	290	−500
5	Cold	10.00	0.8	8.0	90	190	−800
6	Cold	4.00	1.0	4.0	40	190	−600
Total							−50

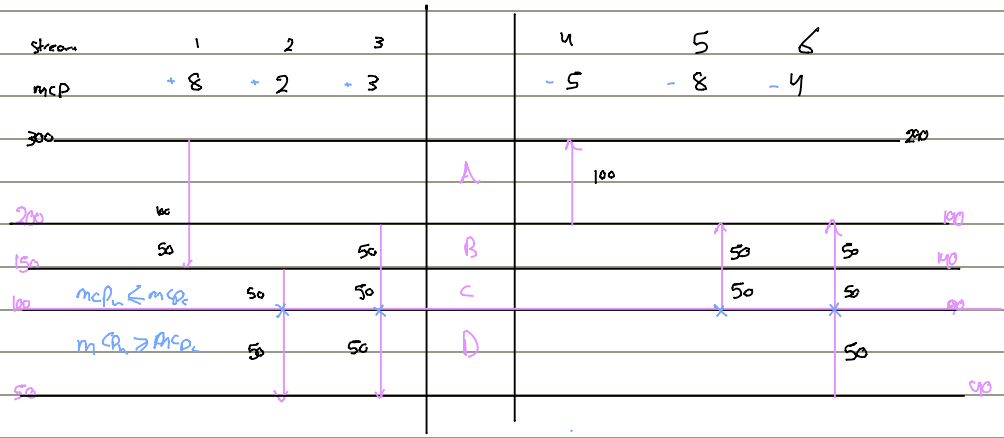


A 100(8−5) = 300

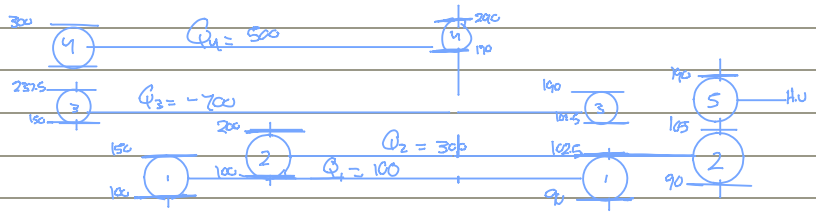
B 50(8+3−8−4) = −50

C 50(2+3−8−4) = −350

D 50(2+3−4) = 50



Above the pinch



First we start in point C

$Q_1 = 2(150 - 100) = 100$

$T_{out} = Q_1 = 100 = -8(90 - T_{out})$

$Q_2 = 3(200 - 100) = 300$

$T_{out} = Q_2 = 300 = -4(90 - T_{out})$

$Q_3 = 8(107.5 - 190) = -700$

$T_{in} = Q_3 = -700 = -8(T_{in} - 150)$

$Q_4 = 4(165 - 190) = -100$

$Q_4 = 8(300 - 287.5) = 500$

$T_{out} = 500 = -5(190 - T_{out})$

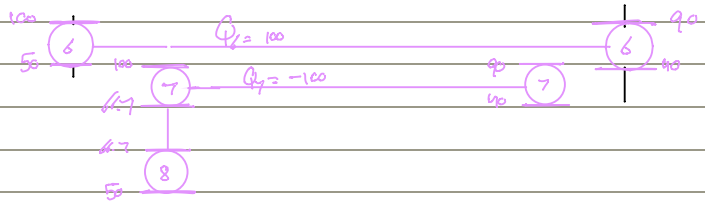
$Q_1 = 2(100 - 50) = 100$

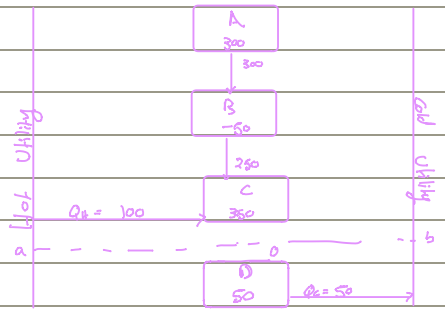
$T_{in} = 100 = -2(T_{in} - 90)$

$Q_7 = 2(40 - 90) = -100$ KW

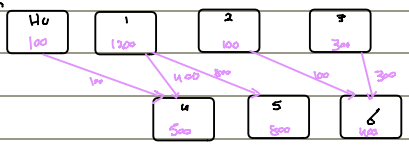
$T_{out} = -100 = -2(100 - T_{out})$

Below the pinch





Above the pinch



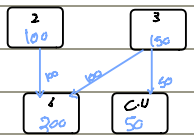
1 $Q = mcp\Delta T = 8\ (150) = 1200$ $N_{min} = 5$

2 $Q = mcp\Delta T = 50 \times 2 = 100$

3 $100(3) = 300$

4. 100×5

bellow the pinch



$N_{min} = 3$

2. What are the advantages and disadvantages of decreasing the minimum approach temperature in a heat-exchange network?

15.2 By decreasing the ΔT_{min} in a HEN, the process-process heat exchangers at the pinch will require larger areas and will therefore be more expensive. However, the amount and cost of hot and cold utilities for the HEN will be reduced.

Problems

For a process, the following process streams must be cooled or heated.

Stream No.	$\dot{m}C_p$ $(10^3 BTU)/hr°F$	Temperature In °F	Temperature Out °F
1	4	600	320
2	4	470	280
3	3	340	580
4	5	300	480

11. Use the MUMNE algorithm for heat-exchanger networks and a minimum approach temperature of 20°F.

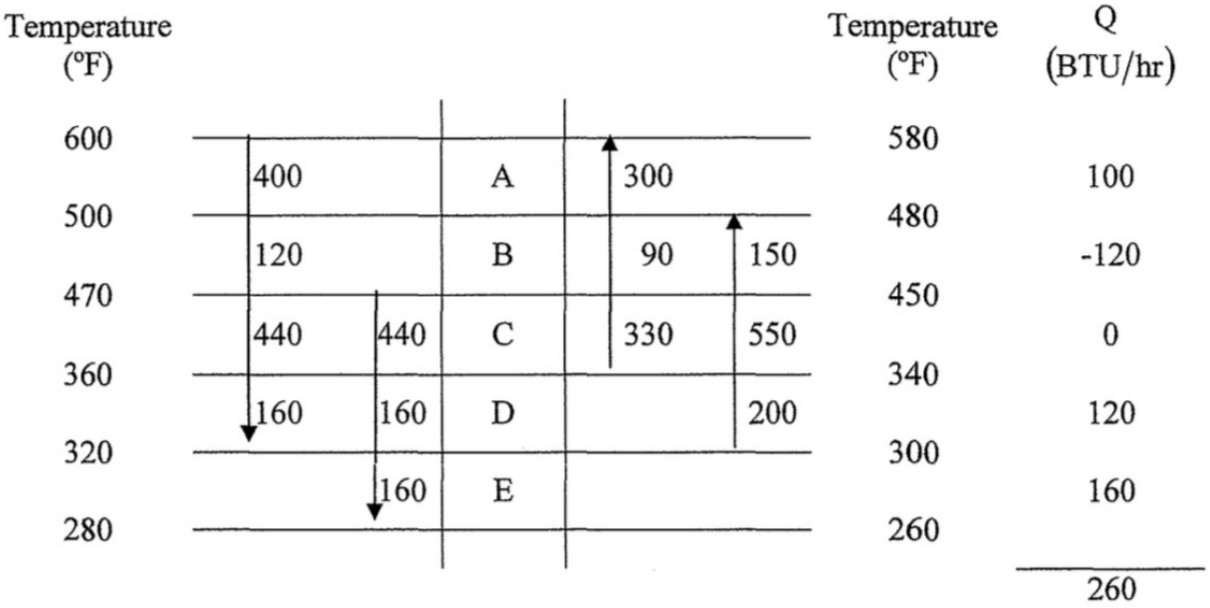
a. Determine the temperature interval diagram.
b. Determine the cascade diagram, the pinch temperatures, and the minimum hot and cold utilities. If there is a choice, for the sake of uniformity, choose the larger values for the pinch temperatures.
c. Determine the minimum number of heat exchangers above and below the pinch.
d. Determine the heat-exchange network above the pinch.
e. Determine the heat-exchange network below the pinch.

For a new process, the following process streams must be cooled or heated.

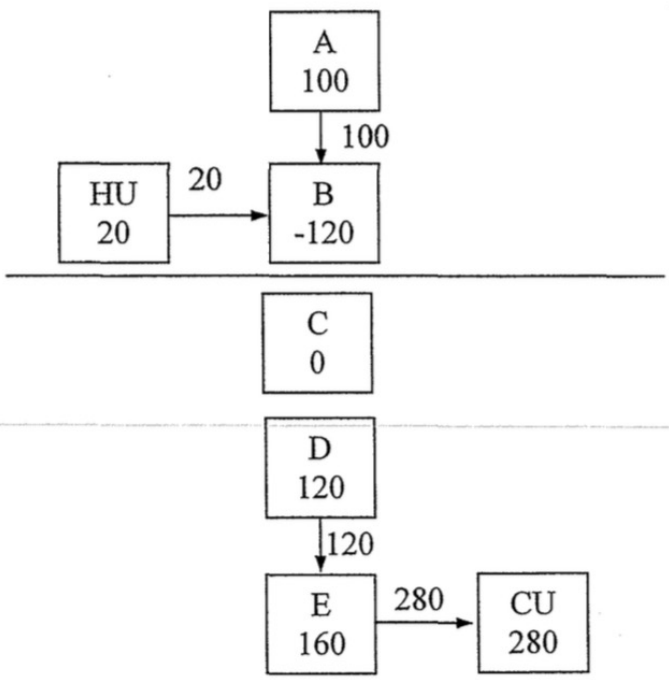
Stream No.	$\dot{m}C_p$ $(10^3 BTU)/hr°F$	Temperature In °F	Temperature Out °F
1	2	400	320
2	4	300	100
3	3	90	310
4	2	170	310

15.11 (a)

$\dot{m}C_P$ (BTU/hr°F)	4	4	3	5
Stream Number	1	2	3	4



(b)

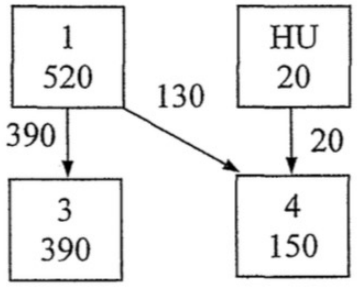


Pinch at 470-450 °F

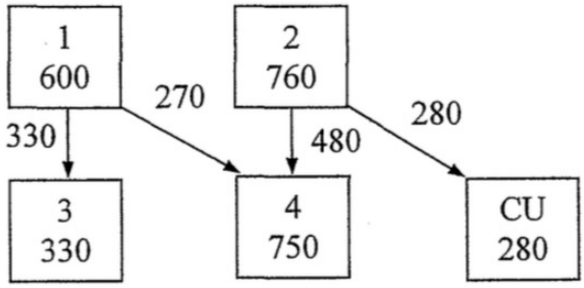
HU = 20 BTU/hr
CU = 280 BTU/hr

(c)

Above
3 exchangers

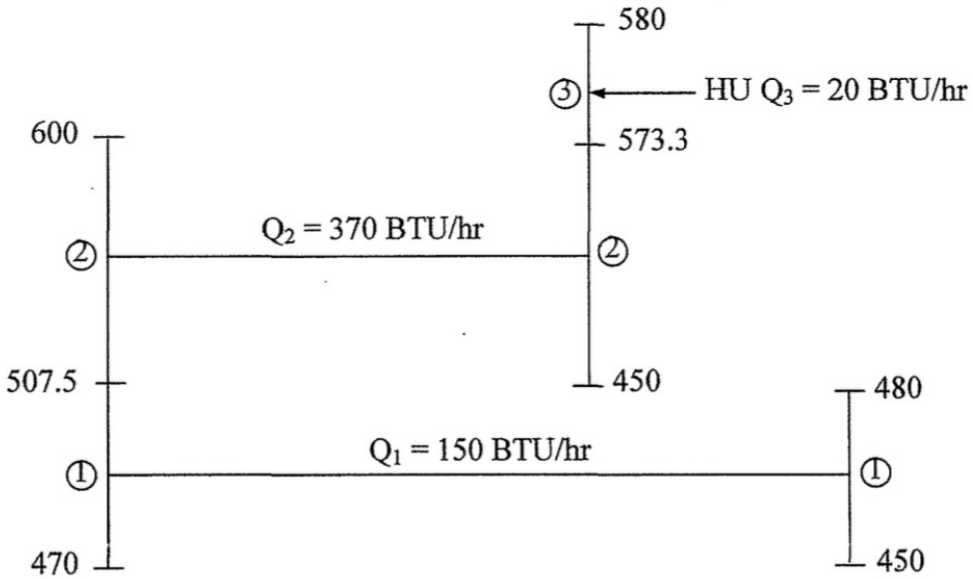


Below
4 exchangers



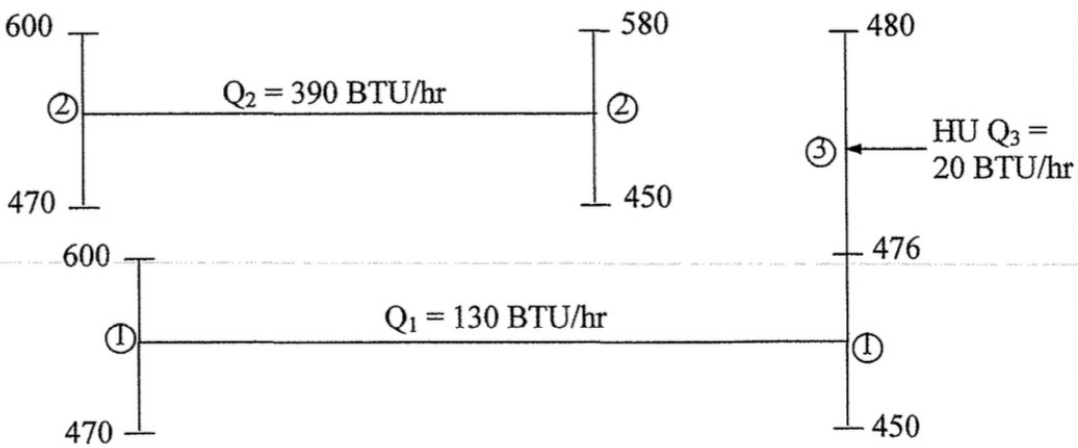
(d) Above $\dot{m}C_{PH} \leq \dot{m}C_{PC} \Rightarrow$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	5



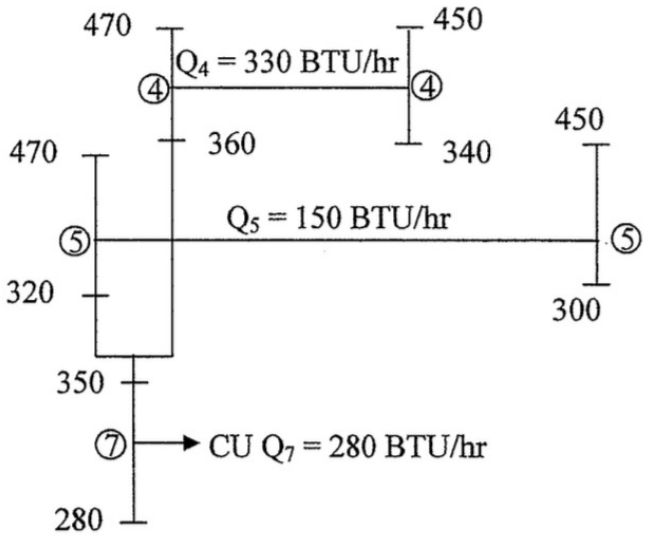
Or split stream 1

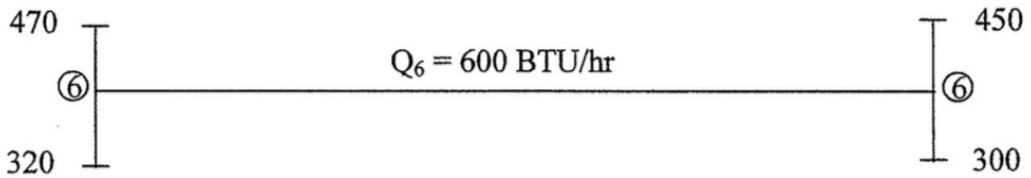
Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	5
	3 1			



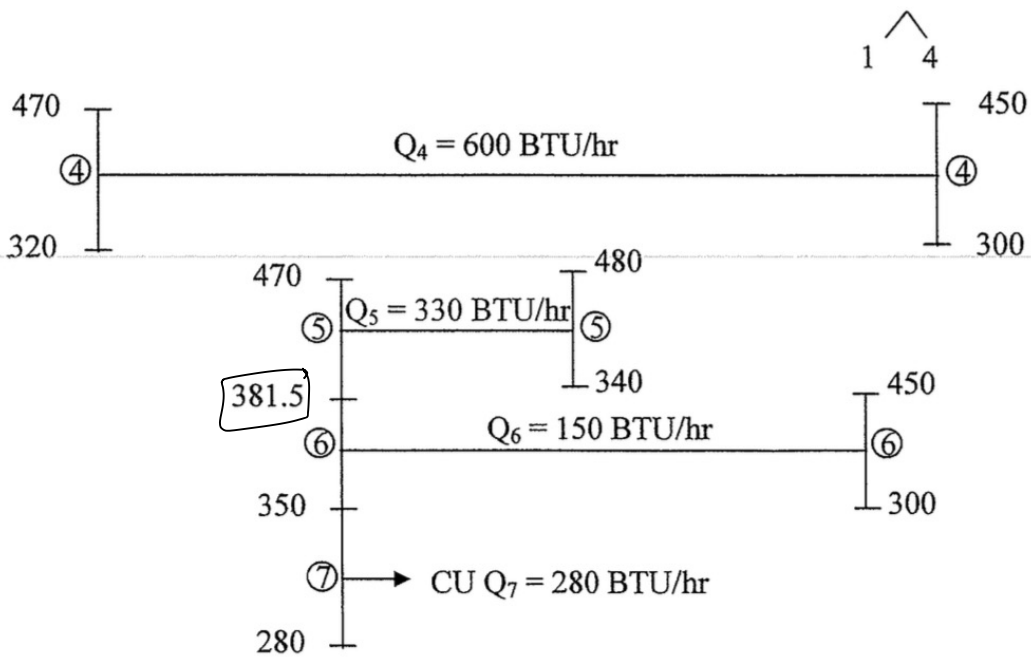
(e) Below pinch $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	5
		1 3		1 4

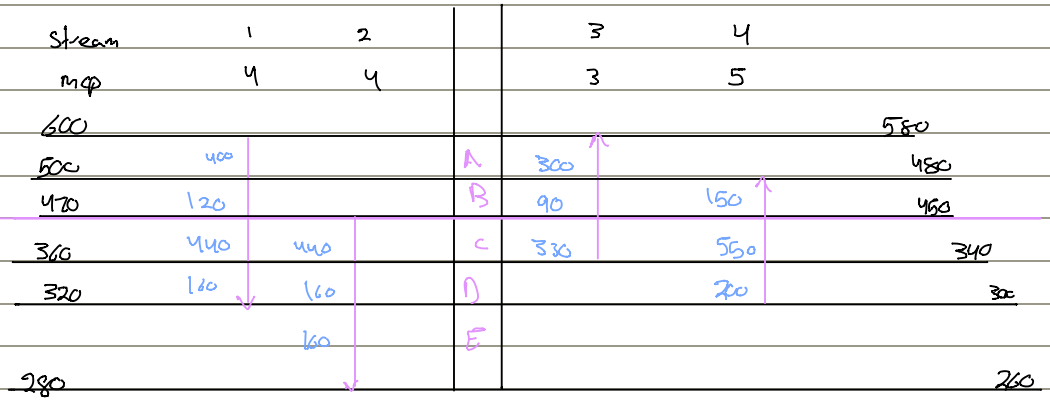




OR



a)



$\Delta T = 20$

$A = 100(4-3) = 100$

$B = 30(4-3-5) = -120$

$C = 110(4+4-3-5) = 0$

$D = 40(4+4-5) = 120$

$E = 40(4) = 160$

1 ↔ 3

$Q_1 = 1(600-470) = 130$

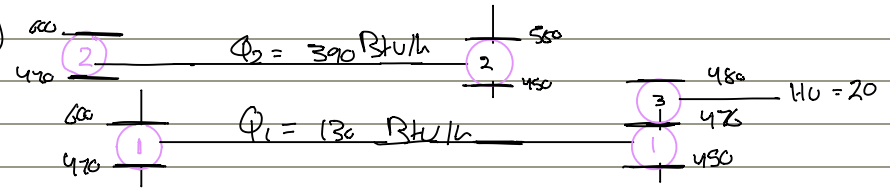
$130 = -5(450 - T_{out})$

$T_{out} = 476$

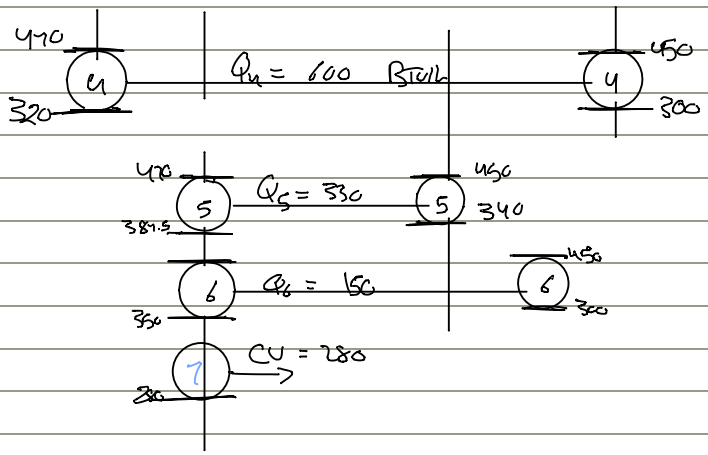
$Q_2 = 3(600 - 470) = 390$

$390 = -3(450 - T_{out})$

d)



e)



4 4

3 5 ↑ 4

$\dot{m} C_p \geq \dot{m} C_p$

$4 > 1$ ✓ $4 > 3$

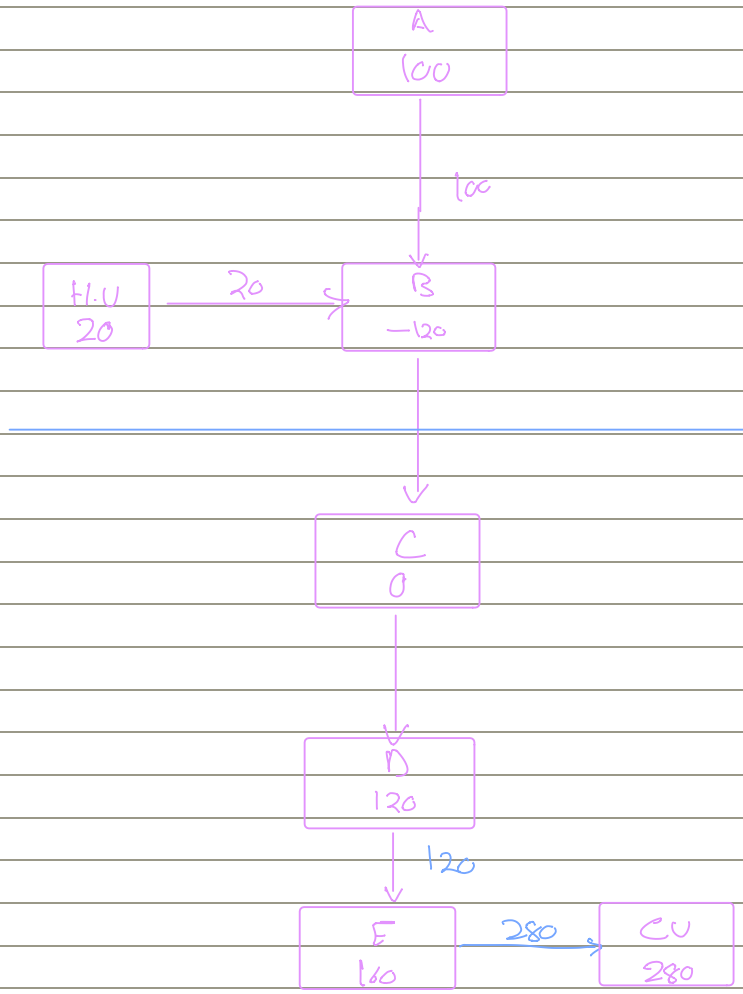
$4 \geq 4$ $4 > 3$

$Q_4 = (450 - 350)$

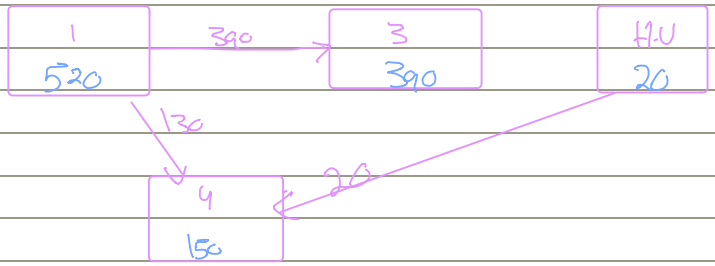
$Q_5 = 3(450 - 340)$

$330 = -4(T_{in} - 476)$

b)



c)



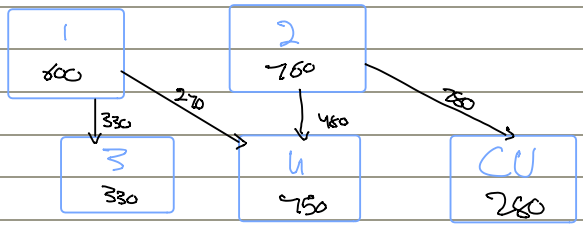
600 - 470 = 130 x 4

580 - 450 = 130 x 3

450 - 450 = 30 x 5

$N_{min} = 3$

below the pinch



4 (470 - 320) = 600

4 (470 - 280) = 760

3 (450 - 340) = 330

5 (450 - 300) =

$N_{min} = 4$

12. Use the MUMNE algorithm for heat-exchanger networks and a minimum approach temperature of 10°F.

a. Determine the temperature interval diagram.
b. Determine the cascade diagram, the pinch temperatures, and the minimum hot and cold utilities.
c. Determine the minimum number of heat exchangers above and below the pinch.
d. Determine the heat-exchange network above the pinch.
e. Determine the heat-exchange network below the pinch.

For a new process, the following process streams must be cooled or heated.

Stream No.	$\dot{m}C_p$ kW/°C	Temperature In °C	Temperature Out °C
1	3	180	100
2	5	120	80
3	3	70	140
4	2	80	160

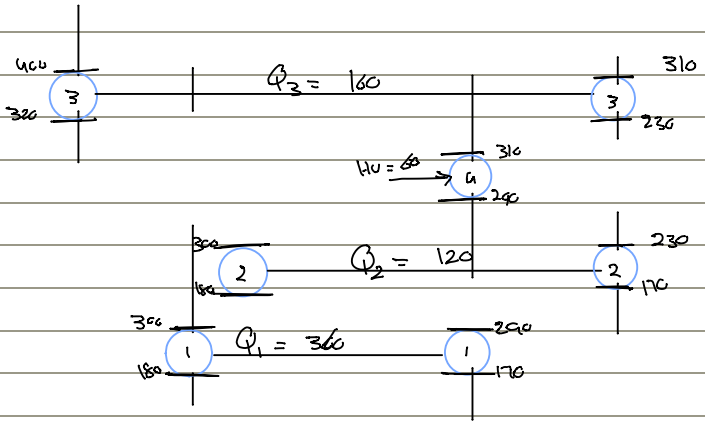
13. Use the MUMNE algorithm for heat-exchanger networks and a minimum approach temperature of 20°C.

a. Determine the temperature interval diagram.
b. Determine the cascade diagram, the pinch temperatures, and the minimum hot and cold utilities.
c. Determine the minimum number of heat exchangers above and below the pinch.
d. Determine the heat-exchange network above the pinch.
e. Determine a heat-exchange network below the pinch that has only one utility exchanger.

For a new process, the following process streams must be cooled or heated.

12. a)

d)



2 → 4 → 3 → 1

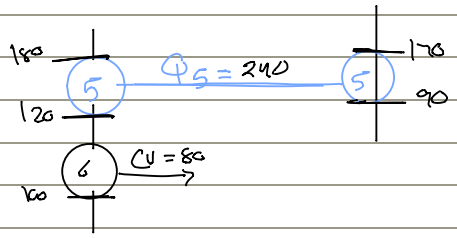
2 → 3

2 → 4

$170 = -2(170 - x)$

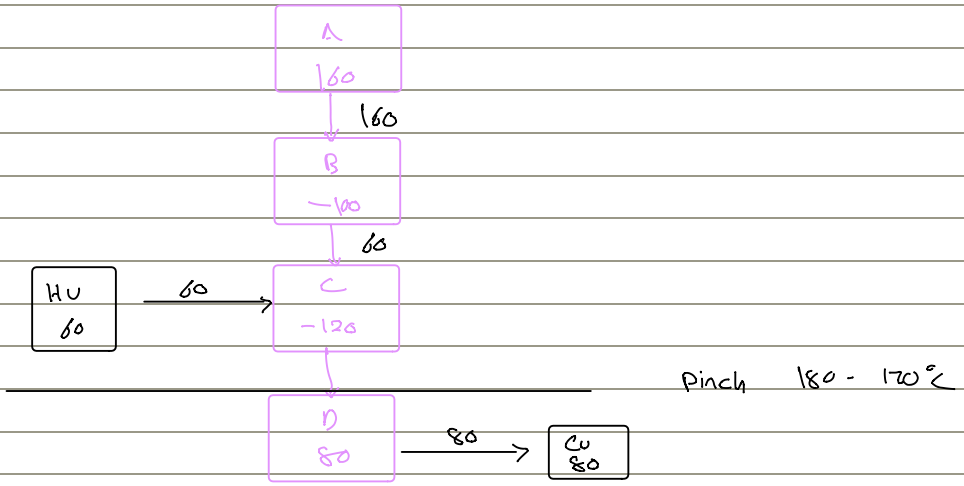
$160 = -2(230 - x)$

e)

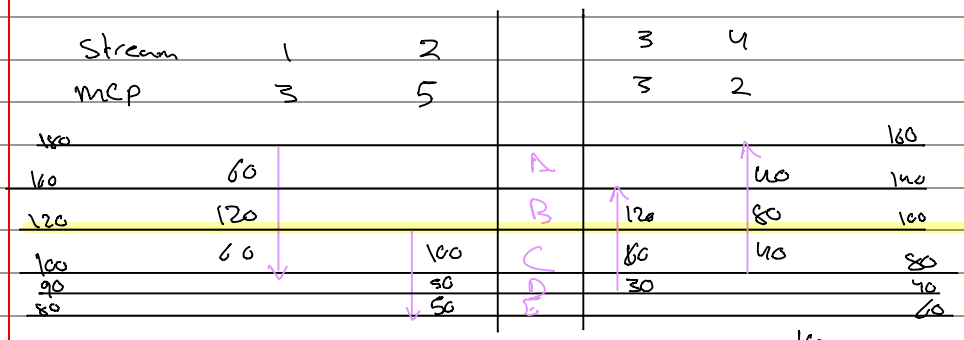


$240 = -4(x - 180)$

b)



13. a



A = 20 (3 - 2)

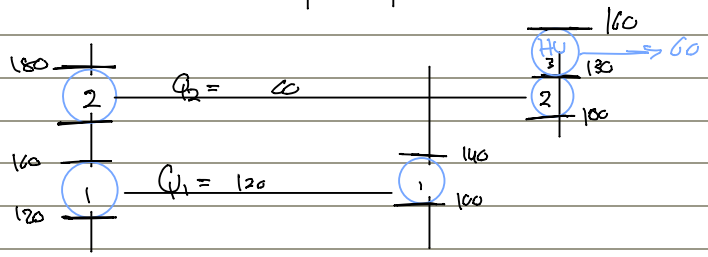
B = 40 (3 - 3 - 2)

C = 20 (3 + 5 - 3 - 2)

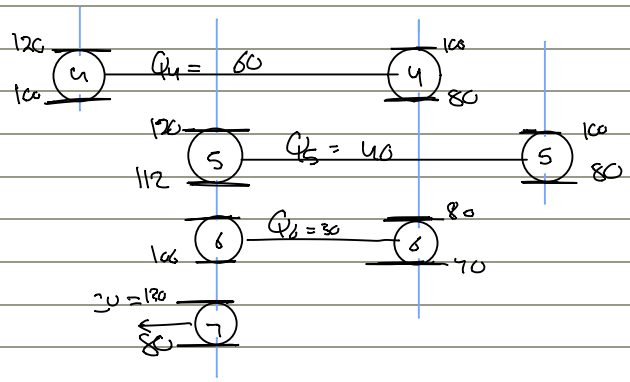
D = 10 (5 - 3)

E = 10 (5)

d)



e)



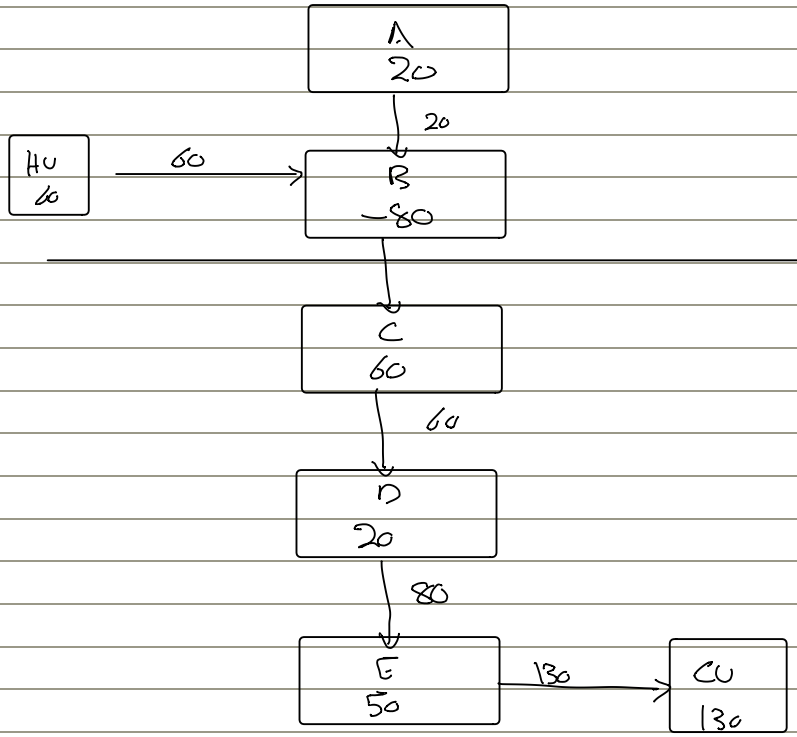
120 = −3 (100 − x)

60 = −2 (100 − x)

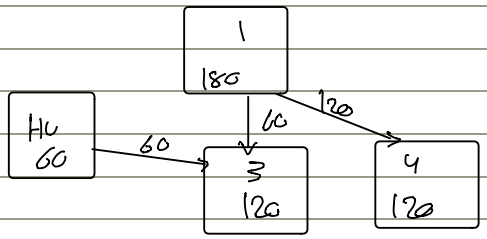
40 = −5 (T − 120)

30 = − 5 (T − 112)

b)



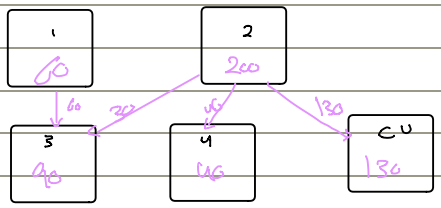
c)



Above the pinch

$N_{min} = 3$

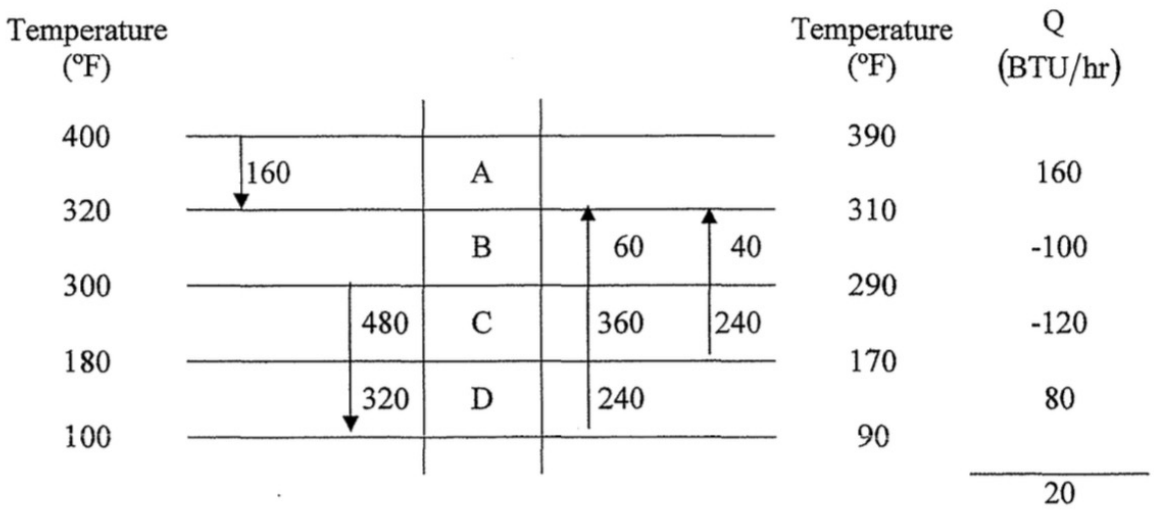
below the pinch



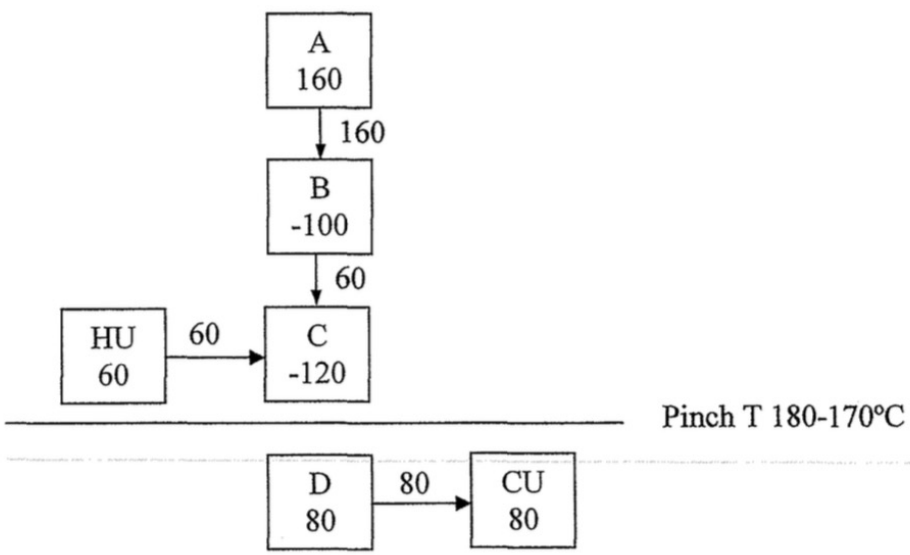
$N_{min} = 4$

15.12 (a)

Stream Number	1	2	3	4
$\dot{m}C_p$ (BTU/hr°F)	2	4	3	2

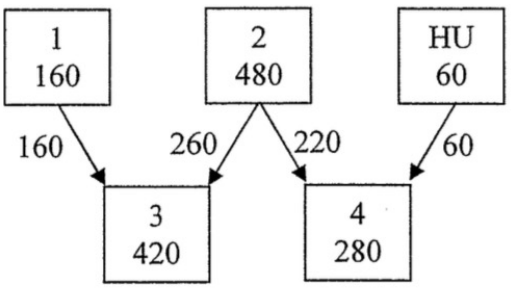


(b)

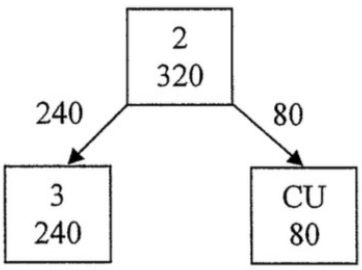


(c)

Above
4 Exchangers

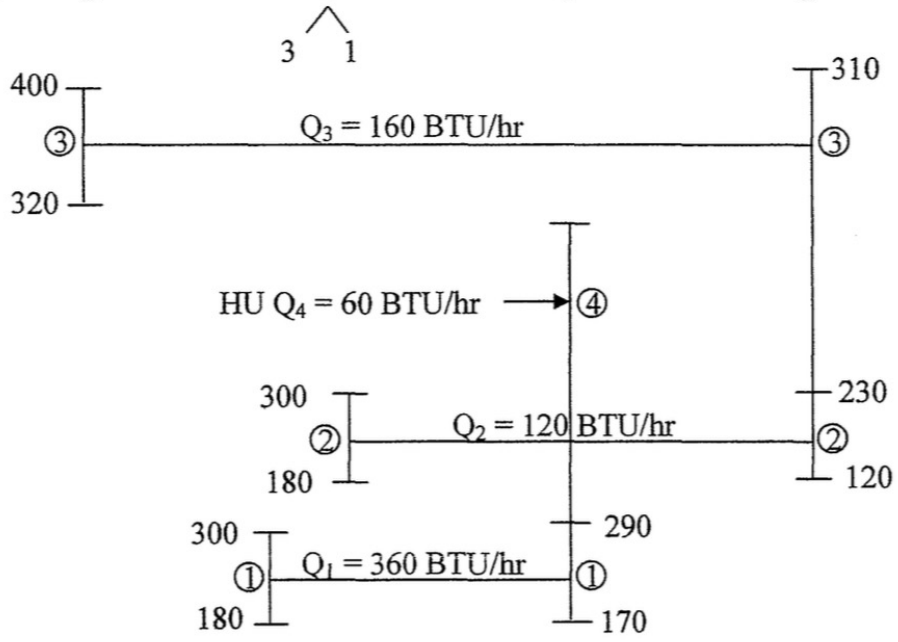


Below
2 Exchangers



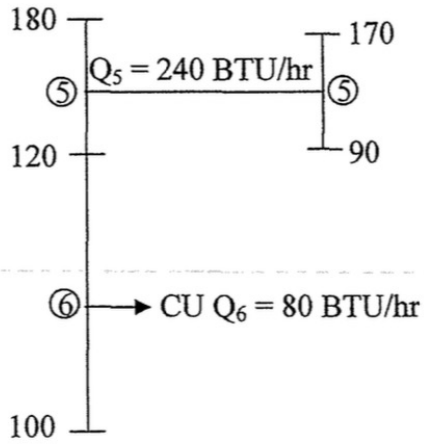
(d) Above pinch $\dot{m}C_{PH} \leq \dot{m}C_{PC}$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	2	4	3	2



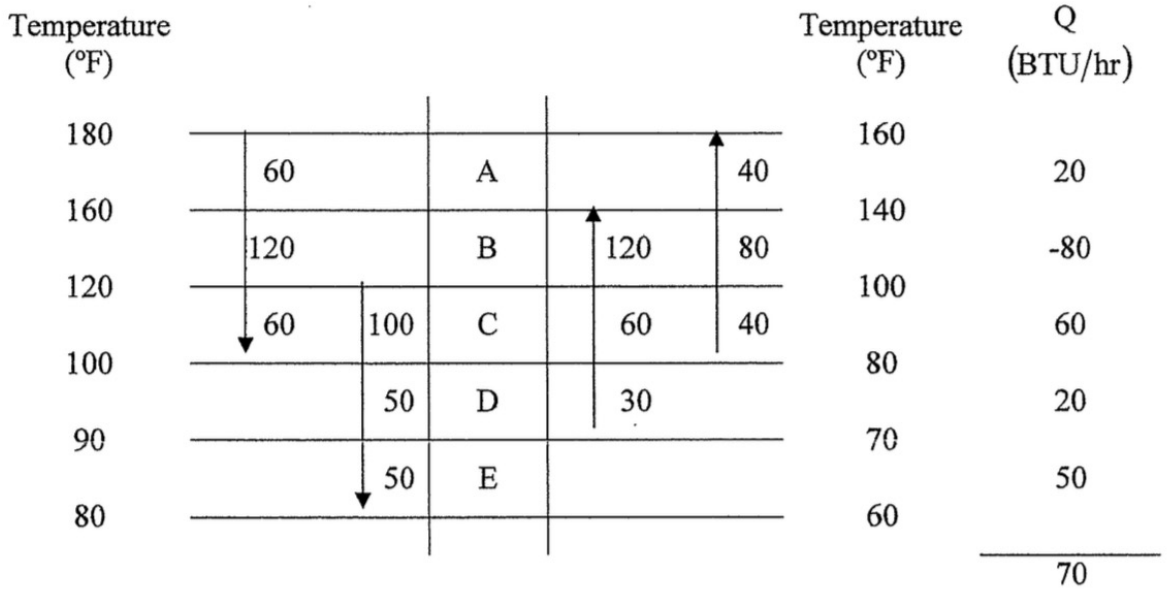
(e) Below pinch $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	2	4	3	2

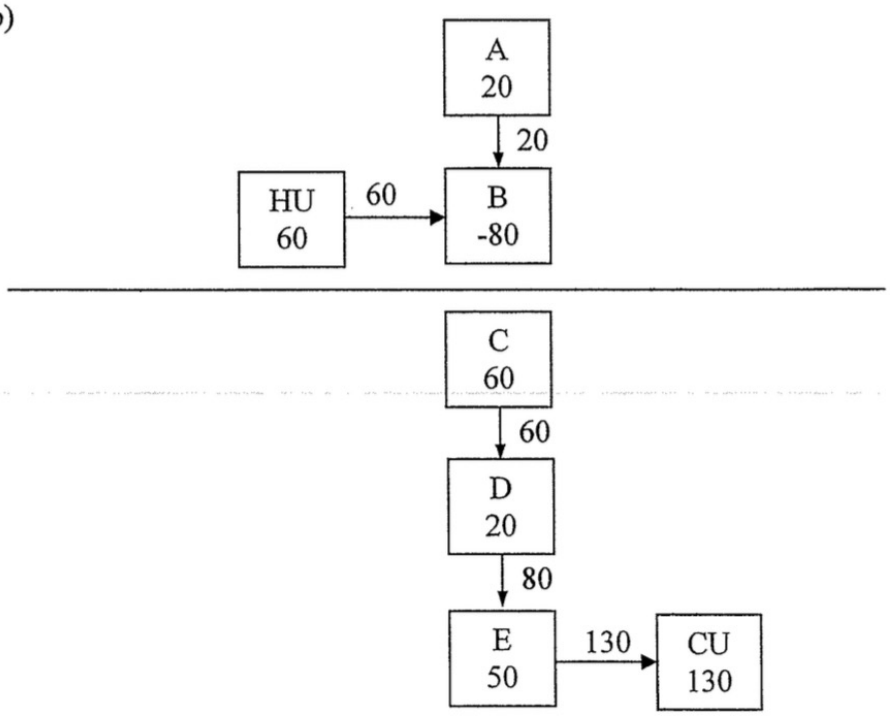


15.13 (a)

Stream Number	1	2	3	4
$\dot{m}C_p$ (BTU/hr°F)	3	5	3	2

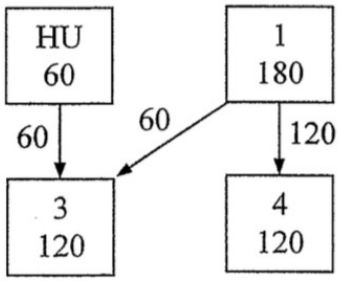


(b)

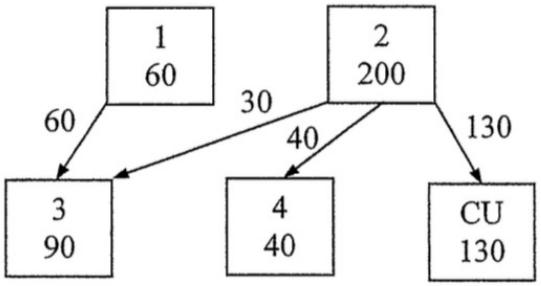


(c)

Above
3 exchangers

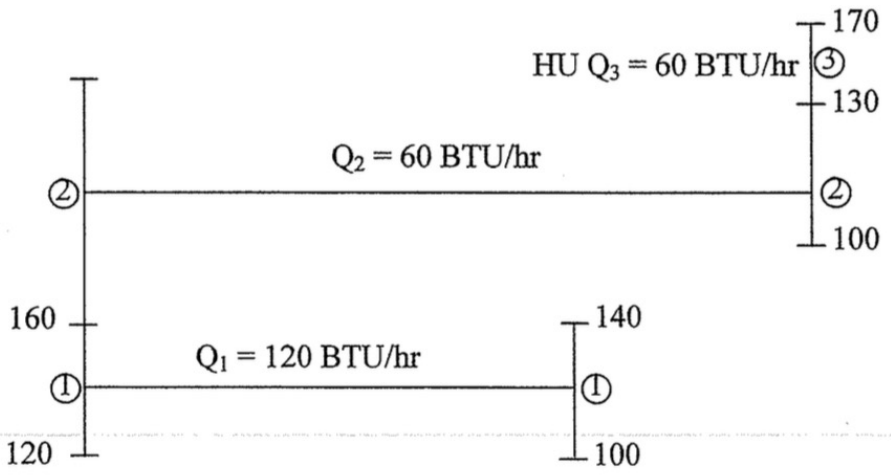


Below
4 exchangers



(d) Above pinch $\dot{m}C_{PH} \leq \dot{m}C_{PC}$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	3	5	3	2

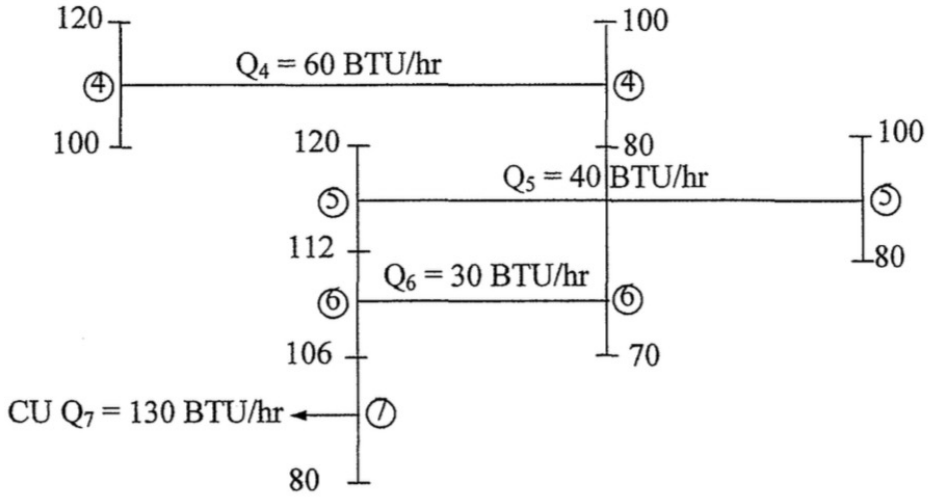


(e)

Below pinch $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

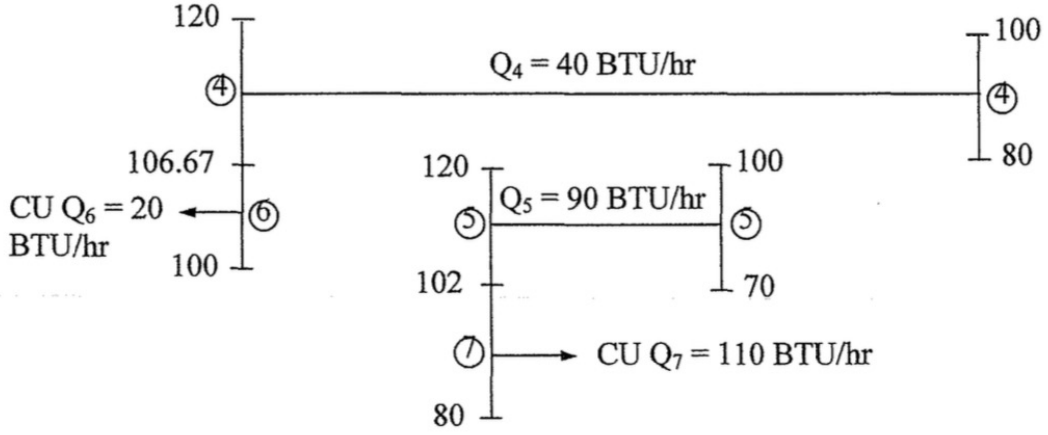
Only one utility stream

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	3	5	3	2



For two utility streams

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	3	5	3	2



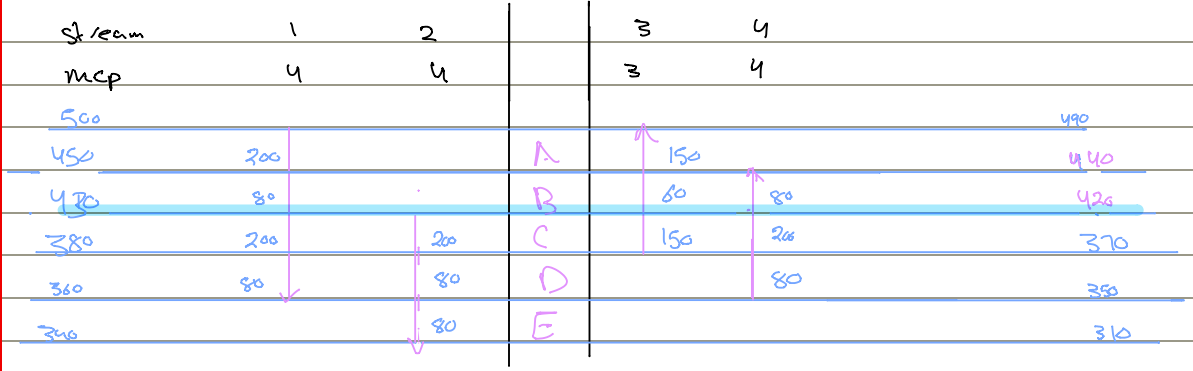
Stream No.	$\dot{m}C_p$ $(10^3 BTU)/hr°F$	Temperature In °F	Temperature Out °F
1	4	500	360
2	4	430	340
3	3	370	490
4	4	350	440

14. Use the MUMNE algorithm for heat-exchanger networks and a minimum approach temperature of 10°F.

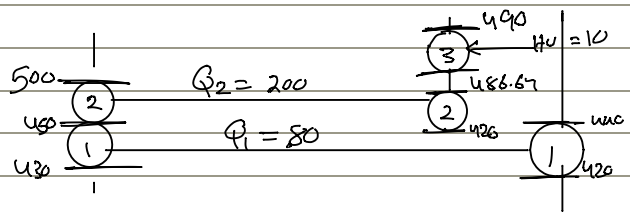
a. Determine the temperature interval diagram.
b. Determine the cascade diagram, the pinch temperatures, and the minimum hot and cold utilities.
c. Determine the minimum number of heat exchangers above and below the pinch.
d. Determine the heat-exchange network above the pinch.
e. Determine a heat-exchange network below the pinch.

In a process design, the following process streams must be cooled or heated.

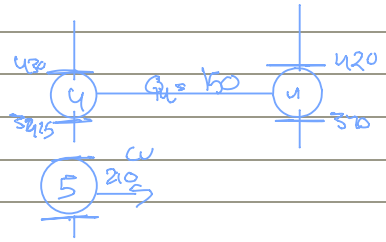
Stream No.	$\dot{m}C_p$ kW/°C	Temperature In °C	Temperature Out °C
1	3	250	200
2	5	200	40
3	4	30	200
4	3	90	200



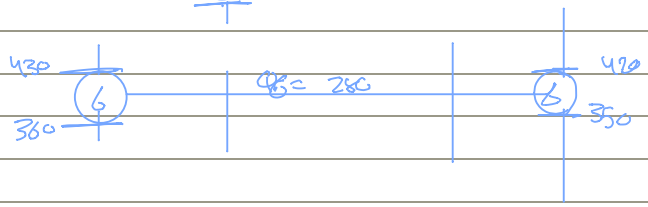
A $50(4-3)$
B $20(4-3-4)$
C $50(4+4-3-4)$
D $20(4+4-4)$
E $20(4)$



$200 = -3(420 - T_{out})$

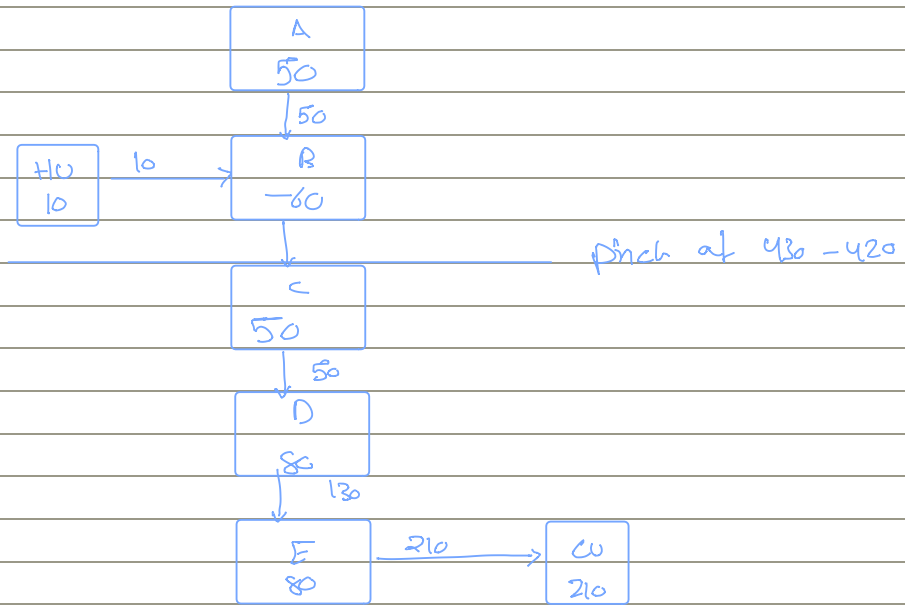


$150 = -4(T_{in} - 430)$

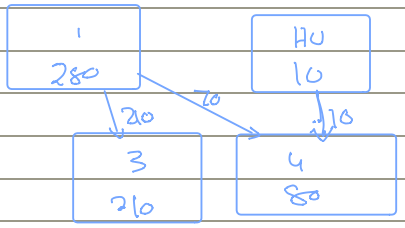


$280 = -4(T_{in} - 420)$

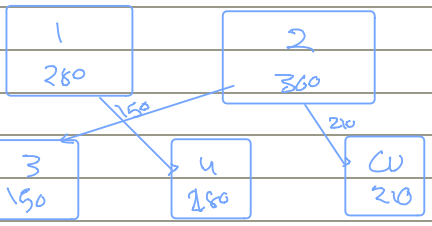
b)



c)



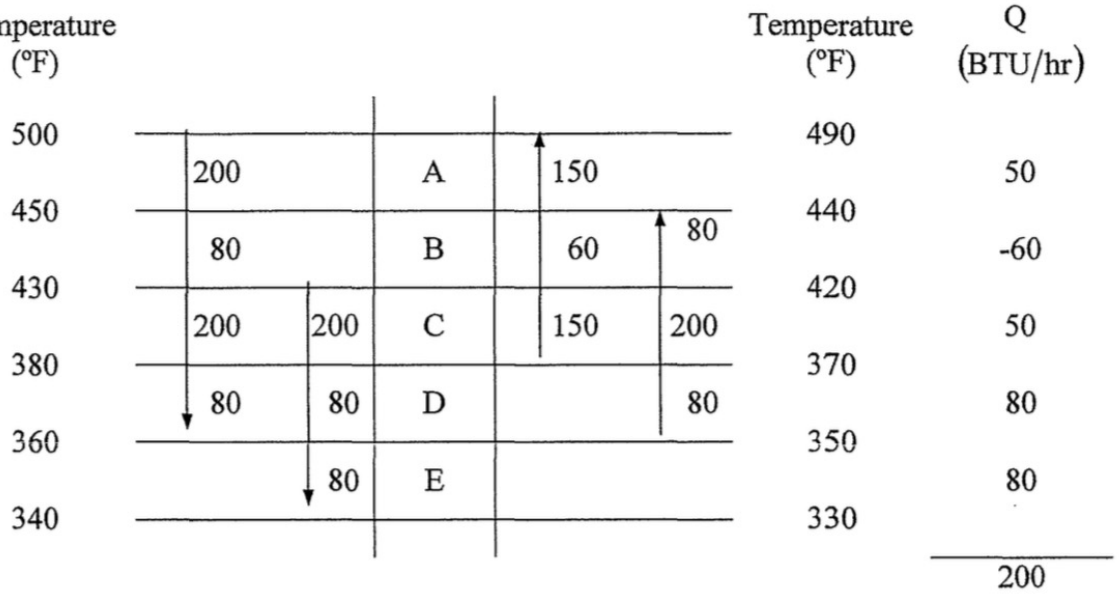
$N_{min} = 3$



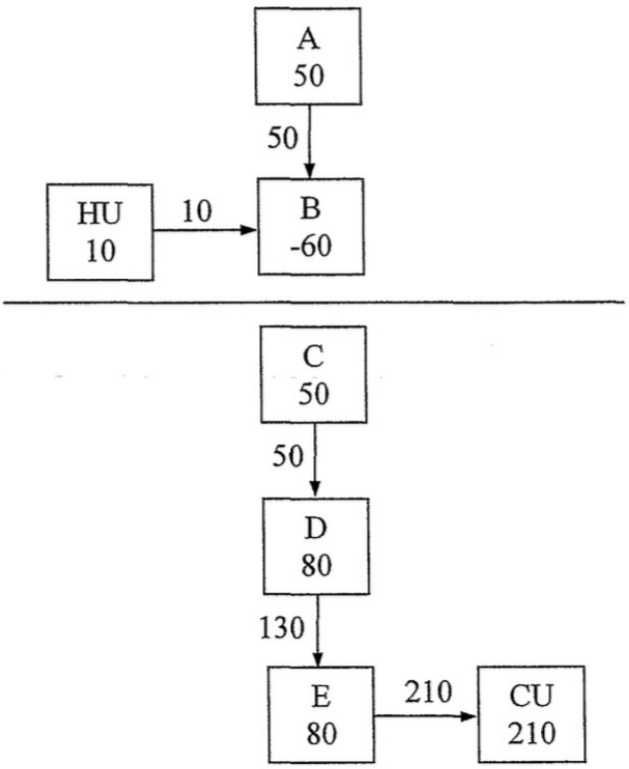
$N_{min} = 3$

15.14 (a)

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	4

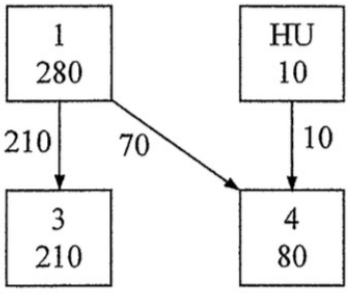


(b)

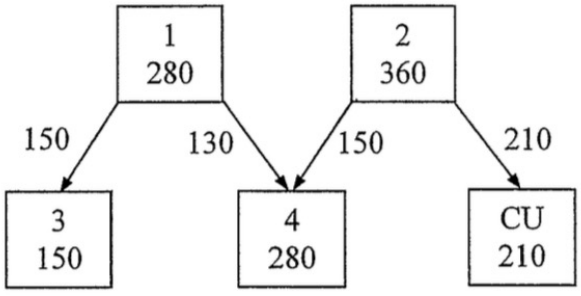


(c)

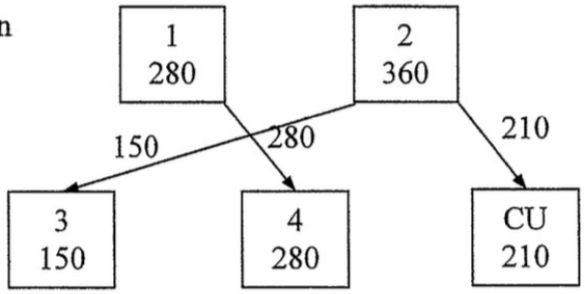
Above pinch
3 exchangers



Below pinch
4 exchangers

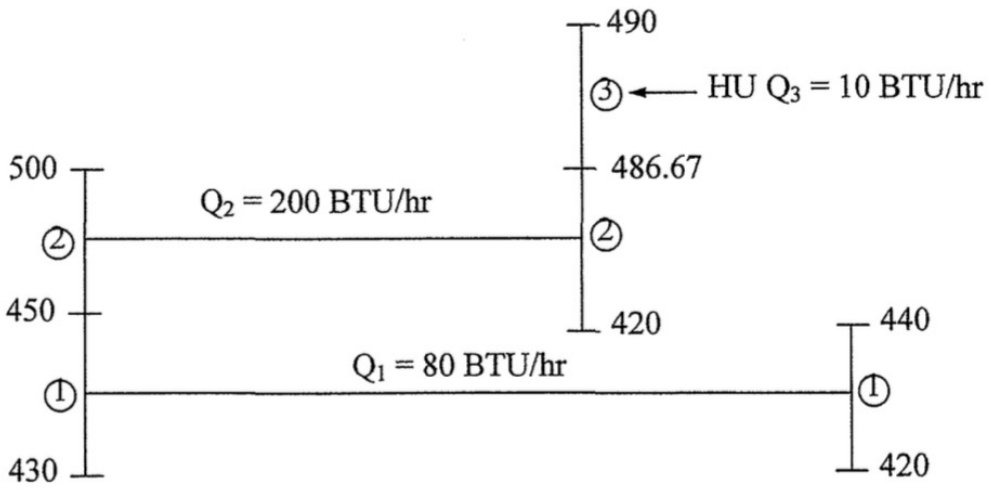


Exact Match Solution
Below pinch
3 exchangers



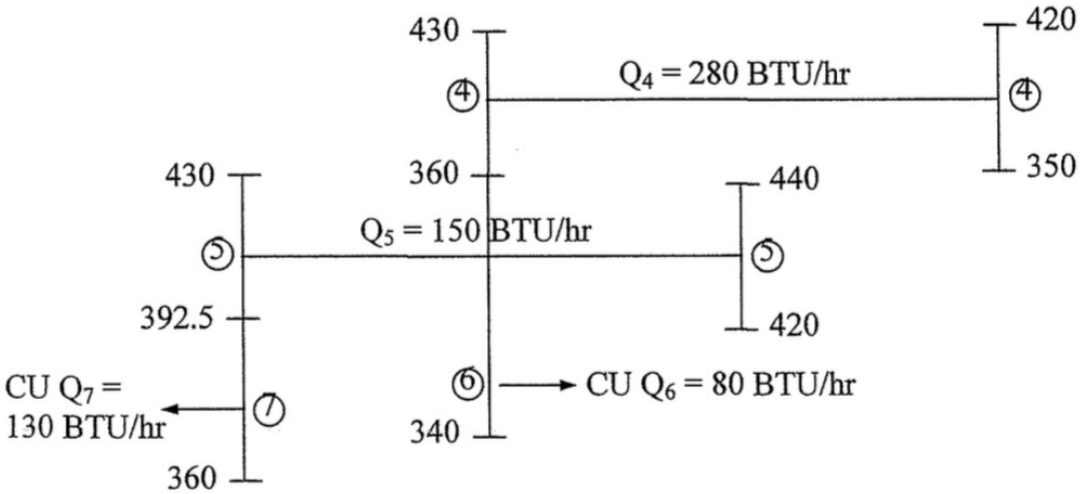
(d) Above pinch $\dot{m}C_{PH} \leq \dot{m}C_{PC}$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	4



(e) Below pinch $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	4



Exact match solution

Stream Number	1	2	3	4
$\dot{m}C_P$ (BTU/hr°F)	4	4	3	4