

UOP ENGINEERING DESIGN SEMINAR

HYDROPROCESSING

Fall 1991

HYDROPROCESSING

I. BACKGROUND

DEFINITIONS

CLASSIFICATIONS

HISTORY

CHEMISTRY *TIPO DE REACCIONES*

KINETICS

II. PROCESS DESIGN *ASPECTOS BASICOS*

GENERAL

MATERIAL BALANCES

YIELD ESTIMATES

MATERIAL (MOL) BALANCES

EXAMPLES

ENERGY BALANCES

FLOW SCHEMES *POSSIBLE INTERCONEXIONES*

CAPITAL VS UTILITY COSTS

FLEXIBILITY AND OPERABILITY

III. FRACTIONATION REQUIREMENTS

PRODUCT QUALITY

PHYSICAL CONSTRAINTS

ECONOMICS

HYDROCRACKING

REDUCIONADO CON

- * CATALYTIC REACTIONS
- * REACTIONS WITH HYDROGEN
- * REDUCE BOILING RANGE

EJ:

NAPHTHA → LPG

EJ:

TRANSFORMATION
VACUUM GAS OIL → NAPHTHA
KEROSENE
DIESEL

- * HYDROTREATS FEEDSTOCK
- * UPGRADES PRODUCTS
- * UNIFORM HIGH QUALITY PRODUCTS CON RESPECTO A LA CARGA

HYDROTREATING

CONVERSION DE INSAT A SATL SATURADO
Y PREPARACION DE CARGAS PARA POSTERIORES
PROCESOS.

- * CATALYTIC $2 \times$
- * REACTIONS WITH HYDROGEN
- * STABILIZES PRODUCTS (ESTABILIZACION DE PRODUCTOS)

CONVERSION OF UNSTATURATES $\longrightarrow$ PARAFFINS

OLEFINS

UNSTABLE DIOLEFINS (GUM FORMING)

- * REMOVES OBJECTIONAL ELEMENTS TÍPICAMENTE ENCONTRADOS
TRANSFORMANDOLOS. COMO *

SULFUR	$\longrightarrow$	HYDROGEN SULFIDE
NITROGEN	$\longrightarrow$	AMMONIA
OXYGEN	$\longrightarrow$	WATER
HALIDES	$\longrightarrow$	HYDROGEN HALIDE
TRACE METALS	$\longrightarrow$	SE DEPOSITAN EN EL CATALIZ CATALYST

- * PRODUCT BOILING RANGE $\approx$ FEED

* PREPARACION DE LA CARGA PARA PROCESOS REFORMACION QUE NO
TOLERAN CONTAMINACION POR ELEMENTOS LISTADOS.

FEED STOCK TYPES

* VIRGIN FEEDS DE DESTIL PRIMARIA

* CRACKED FEEDS

THERMAL VISBREAKING, HIDROCRACKEO

CATALYTIC DE FCC

* DEASPHALTED FEEDS

* SYNTHETIC HYDROCARBONS

VIRGIN STOCKS

SE PURVEN DESINDE COMO:

* NAPHTHAS

* DISTILLATES

KEROSINE

DIESEL

GAS OILS

* VACUUM GAS OIL

* REDUCED CRUDE (^{343°C} 650 F +)

* VACUUM RESID (^{565°C} 1050 F +)

CRACKED FEEDS

* ORIGINADAS EN PROCESOS DE TIPO
THERMAL (VISBREAKING)

NAPHTHAS

PYROLYSIS

COKER, ETC

DISTILLATES

DIESELS

GAS OILS

* **CATALYTIC CRACKED (FCC)**

LIGHT CYCLE OIL RANGO DIESEL

HEAVY CYCLE OIL

HYDROPROCESSING HISTORY

- * UOP'S FIRST INVOLVEMENT WITH HYDROPROCESSING DATES TO THE MID-1930'S
HYDROGENATION OF CB POLYMER (ISOOCETENE) TO ISOOCETANES FOR AVIATION FUEL.
- * UOP'S COMMERCIALIZATION OF PLATFORMING IN 1949 MADE POSSIBLE COMMERCIALIZATION OF HYDROTREATING.
BYPRODUCT HYDROGEN AVAILABLE
REASONABLE PURITY
REASONABLE COST
SUFFICIENT QUANTITIES
- * HYDROPROCESSING IN THE FIFTIES
UPGRADING UNSTABLE MATERIALS (CRACKED PRODUCTS)
NAPHTHA'S
GASOILS
PLATFORMER FEED PRETREATMENT (NAPHTHA HDS)
- * HYDROPROCESSING IN THE SIXTIES
TWO-STAGE HYDROTREATING OF PYROLYSIS GASOLINE
DESULFURIZATION OF DISTILLATE FUELS
UPGRADING KEROSENE FOR JET FUEL
INITIATION OF HIGH PRESSURE HYDROPROCESSING

HYDROPROCESSING HISTORY

* HYDROPROCESSING IN THE SEVENTIES

PRODUCT UPGRADING WITH HYDROCRACKING
REDUCED CRUDE DESULFURIZATION
UPGRADING BOTTOM OF BARREL
ENVIRONMENTAL CONSIDERATIONS

* HYDROPROCESSING IN THE EIGHTIES

SYNTHETIC HYDROCARBON FEEDSTOCKS
IMPROVED ECONOMICS
INCREASED FLEXIBILITY

* IN TERMS OF BARRELS PROCESSED HYDROTREATING IS MOST WIDELY APPLIED PROCESS TECHNOLOGY IN INDUSTRY

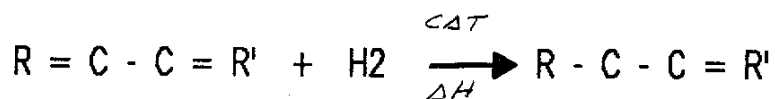
HYDROTREATING REACTIONS

TIPOS DE REACCIONES EN HIDROTRATAMIENTO

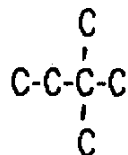
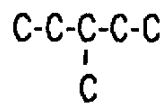
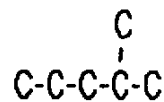
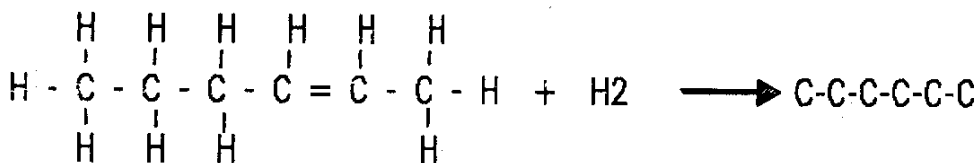
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HYDROGENATION

OLEFIN SATURATION



eg



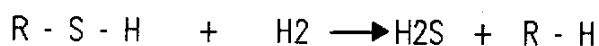
HYDROTREATING REACTIONS

TIPOS DE Rx (CONT...)

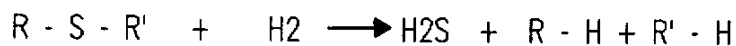
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DESULFURIZATION

MERCAPTANS



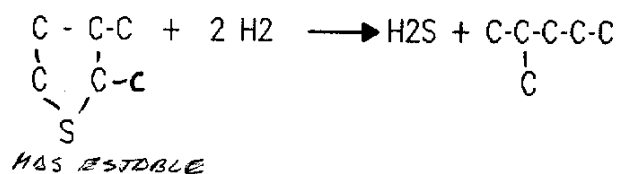
SULFIDES



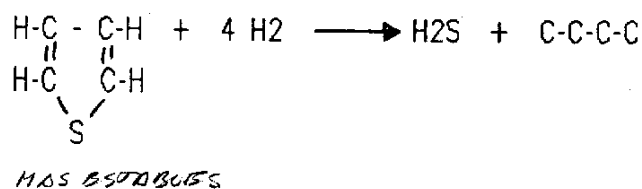
DISULFIDES



CYCLIC SULFIDES



THIOPHENES



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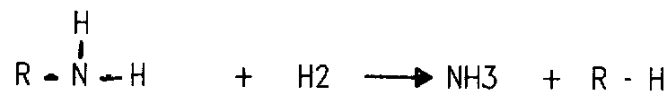
HYDROTREATING REACTIONS

TIROS D12 12x (CONT. -)

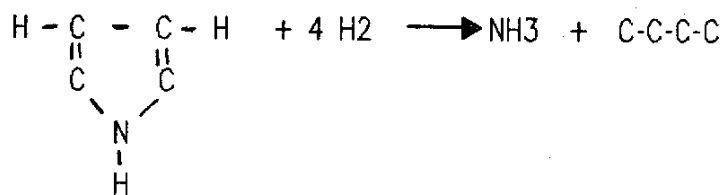
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DENITROGENATION

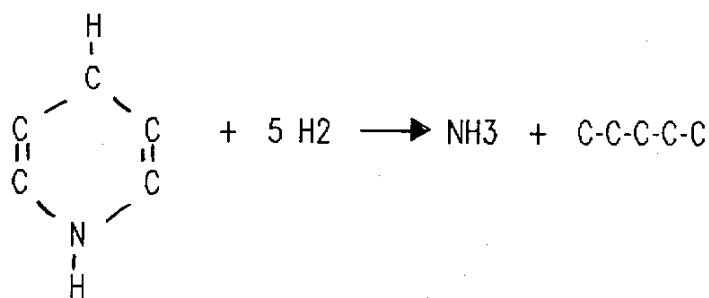
AMINE



PYRROLE



PYRIDINE



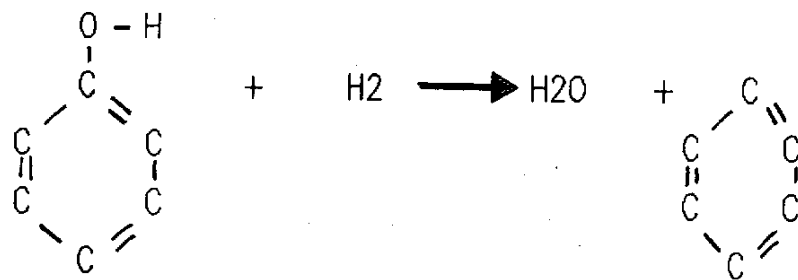
HYDROTREATING REACTIONS

TIPUS DE RX (CONT..)

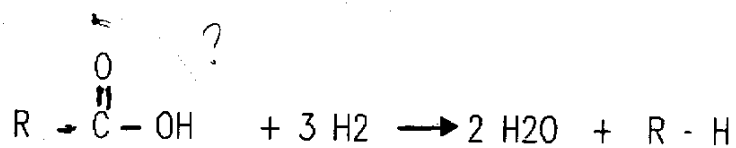
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DEOXIDATION

PHENOLS



PEROXIDES (PROBLEMATICO EN ALMACENAMIENTO)



HYDROTREATING REACTIONS

* DEHALOGENATION

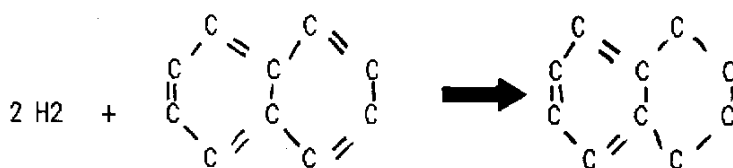
CHLORIDES



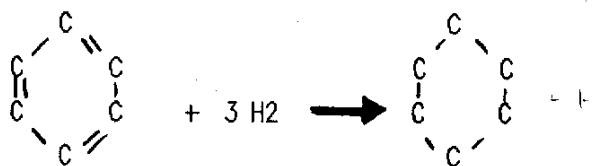
* HYDROGENATION (AROMATICS SATURATION)

IMPORTANTE EN HEDRAMIENTO DE NEROSENE $\rightarrow$ J.F.

POLYAROMATICS



MONOAROMATICS



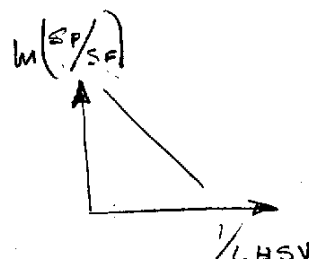
* HYDROCRACKING



HYDROTREATING DESULFURIZATION KINETICS

*

FIRST ORDER REACTIONS



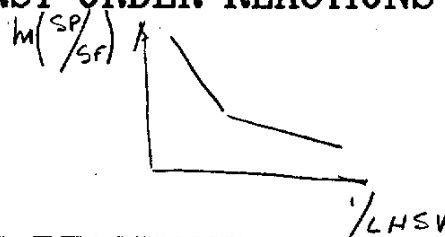
$S = \text{moles de } S$
 $P = \text{PROD}$
 $F = \text{FEED}$

*

TWO SIMULTANEOUS FIRST ORDER REACTIONS

'EASY' SULFUR

'DIFFICULT' SULFUR



*

PSEUDO SECOND ORDER REACTIONS

*

MODERATE DESULFURIZATION LEVELS FOR DISTILLATES AND
 LIGHTER FIRST ORDER KINETIC EXPRESSION ADEQUATELY
 REPRESENTS SYSTEM

*

HIGH SEVERITY OPERATIONS AND FOR HEAVIER OILS (RESIDS)
 EITHER TWO SIMULTANEOUS REACTIONS OR PSEUDO SECOND
 ORDER REACTION CAN BE USED. MOST INVESTIGATORS PERFER
 SECOND ORDER KINETIC EXPRESSION.

DESULFURIZATION KINETICS

FIRST ORDER REACTIONS

- $-r_s$ = rate of disappearance of feed sulfur
 S_F = concentration of sulfur in feed, wt pct
 S_P = concentration of sulfur in product, wt pct
 S = concentration of sulfur in reactor

t = time

τ = space time = $1 / \text{LSHV}$

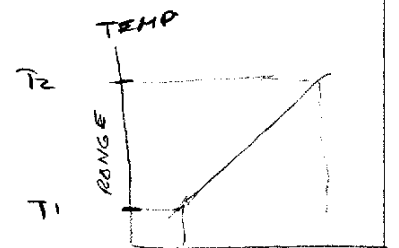
LSHV = Liquid hourly space velocity

$$\text{LSHV} = \frac{\text{FLOW OF REACTANTS [FT}^3/\text{HR}]}{\text{VOLUME OF CATALYST [FT}^3]}$$

$$\text{LSHV} = \frac{1}{\text{HR}} \quad \tau = 1 / \text{LSHV}$$

$$r = f_1(\text{temperature}) \cdot f_2(\text{composition}) = \text{function of } T \text{ \& } \text{COMP}$$

$$r = k \cdot f_2(\text{composition})$$



T_1 : TEMP EST DE STARTUP
 T_2 : TEMP EST DE OPERAC

A MAYOR LHSV MENOR ES
 EL TIEMPO DE RESIDENCIA
 EN EL REACTOR \& MENOR
 CONTACTO CON EL CATALIZADOR.

CONDICIONES

- ① TEMP DE OPERACION DENTRO DE UN RANGO
- ② PRESION DE OPERACION ES FIJA
- ③ VOLUMEN DE CATALIZADOR ES FIJO

k = funcion de temp
 CTE SI TEMP CTE.

$$-r_s = -dS/dt = k S$$

$$\int_{S_F}^{S_P} -dS / S = \int_0^t k dt$$

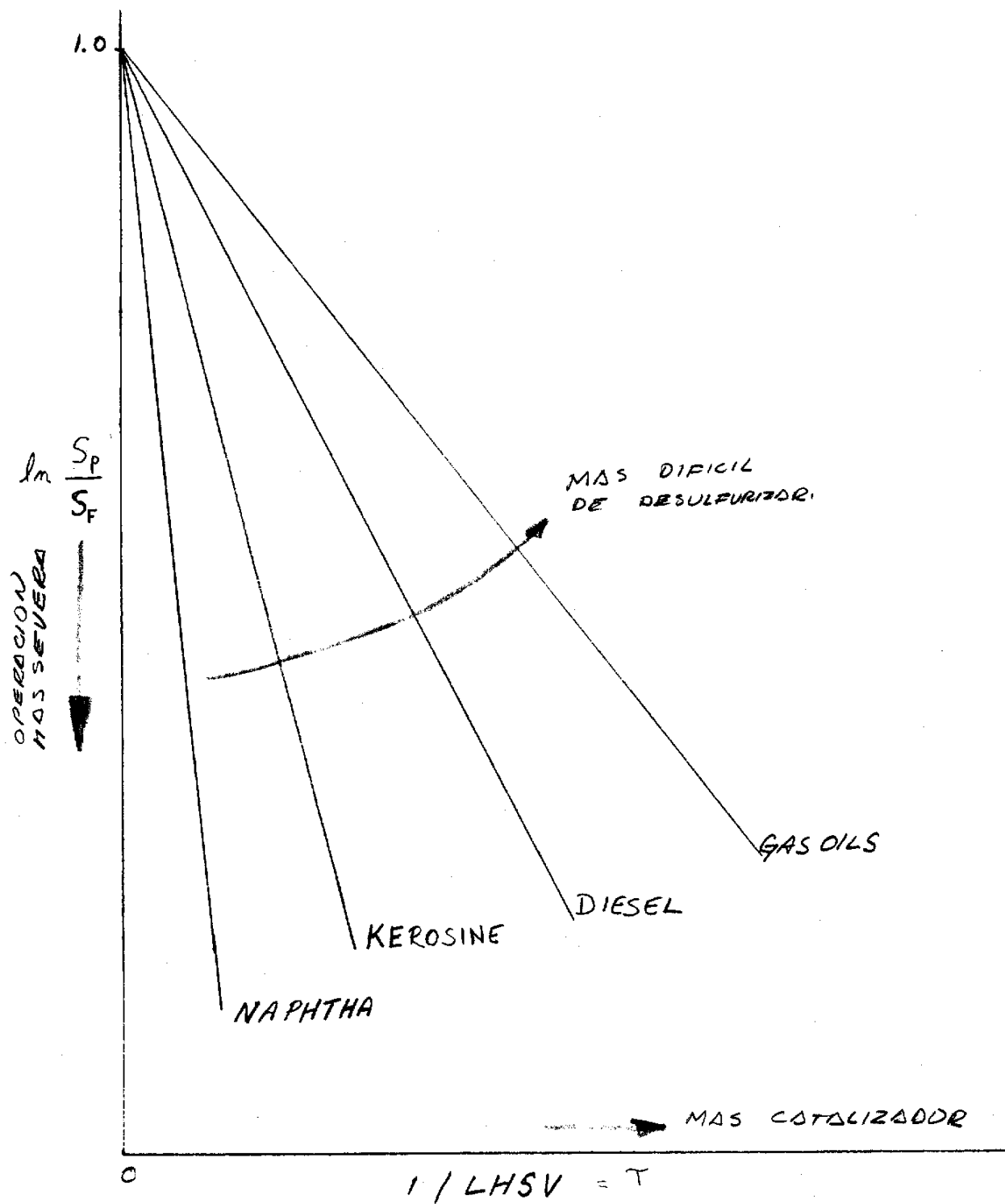
$$-\ln S \Big|_{S_F}^{S_P} = k t$$

$$-\ln S_P + \ln S_F$$

Plug flow reactor $t = \tau$.

$$\ln S_F / S_P = k \cdot \tau = k / \text{LSHV}$$

HYDROTREATING
DESULFURIZATION - FIRST ORDER KINETICS



DESULFURIZATION KINETICS

FIRST ORDER REACTIONS

TEMPERATURE DEPENDENT TERM

ARRHENIUS - TYPE RELATIONSHIP:

$$k = k_0 e^{-E/RT}$$

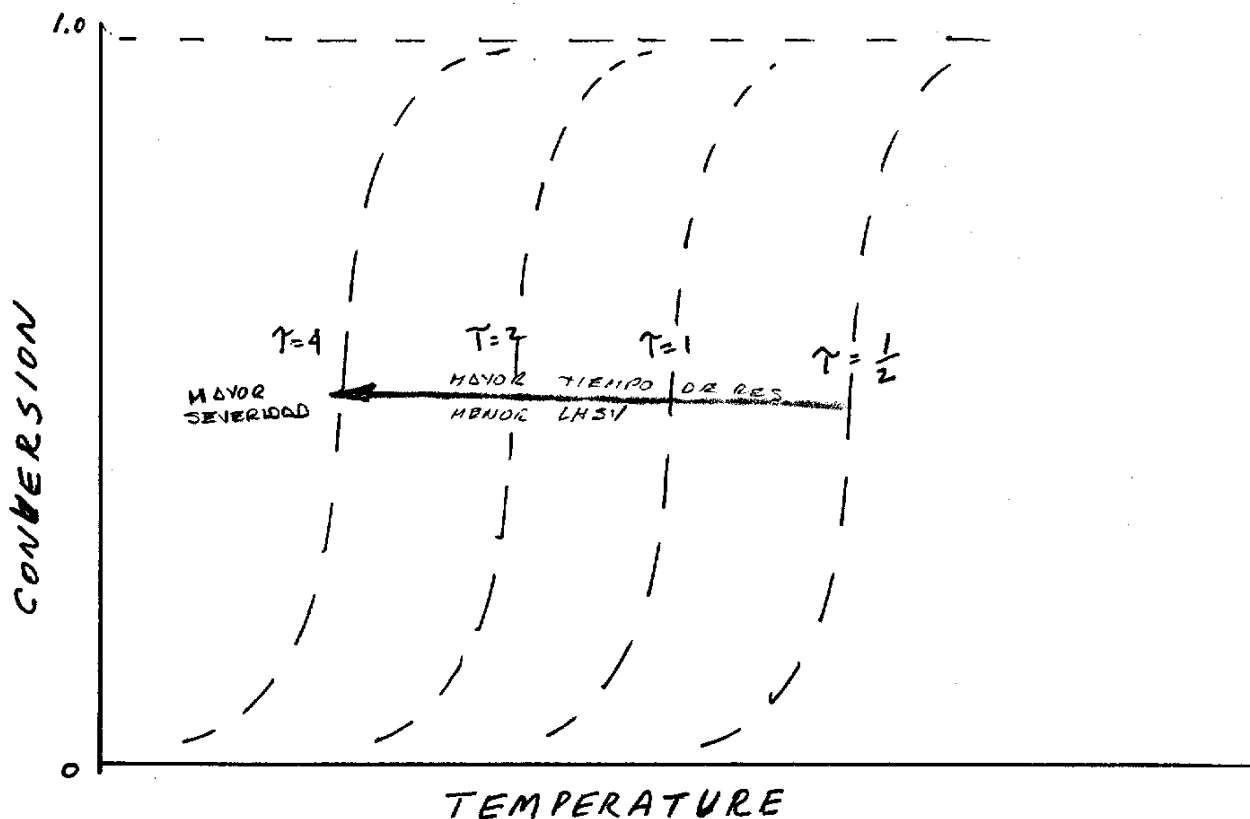
where

E = activation energy

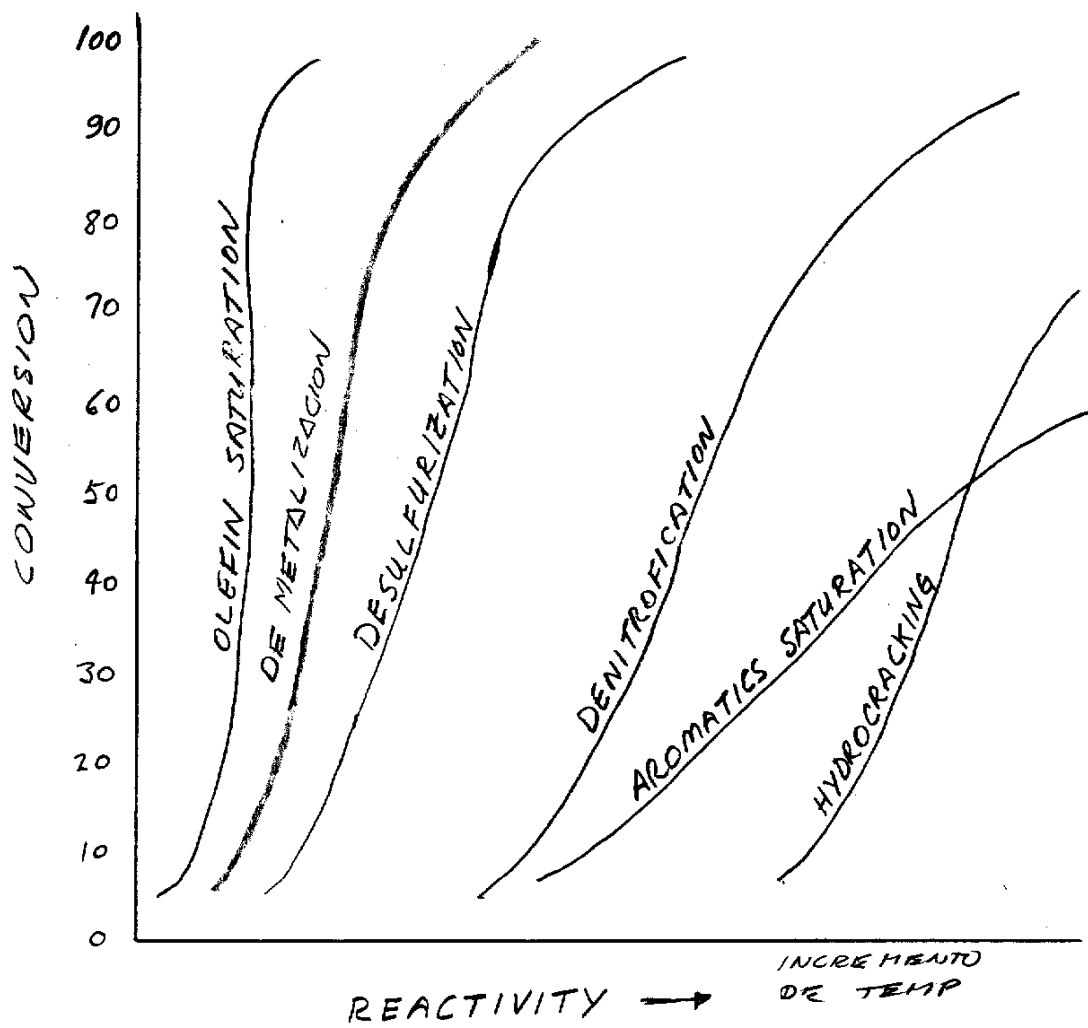
T = temperature

R = gas law constant

for irreversible reactions:



HYDROPROCESSING REACTIONS



HYDROTREATING PROCESS VARIABLES

* PRESSURE

HYDROGEN PARTIAL PRESSURE

CATALYST STABILITY

* TEMPERATURE

* LHSV (SPACE VELOCITY) $\frac{1}{\text{TIME}}$

* HYDROGEN TO FEED RATE (SCFB)

HYDROGEN PARTIAL PRESSURE

* RECYCLE GAS PURITY

INFLUYE EN LA
PRESION PARCIAL
DEL HIDROGENO

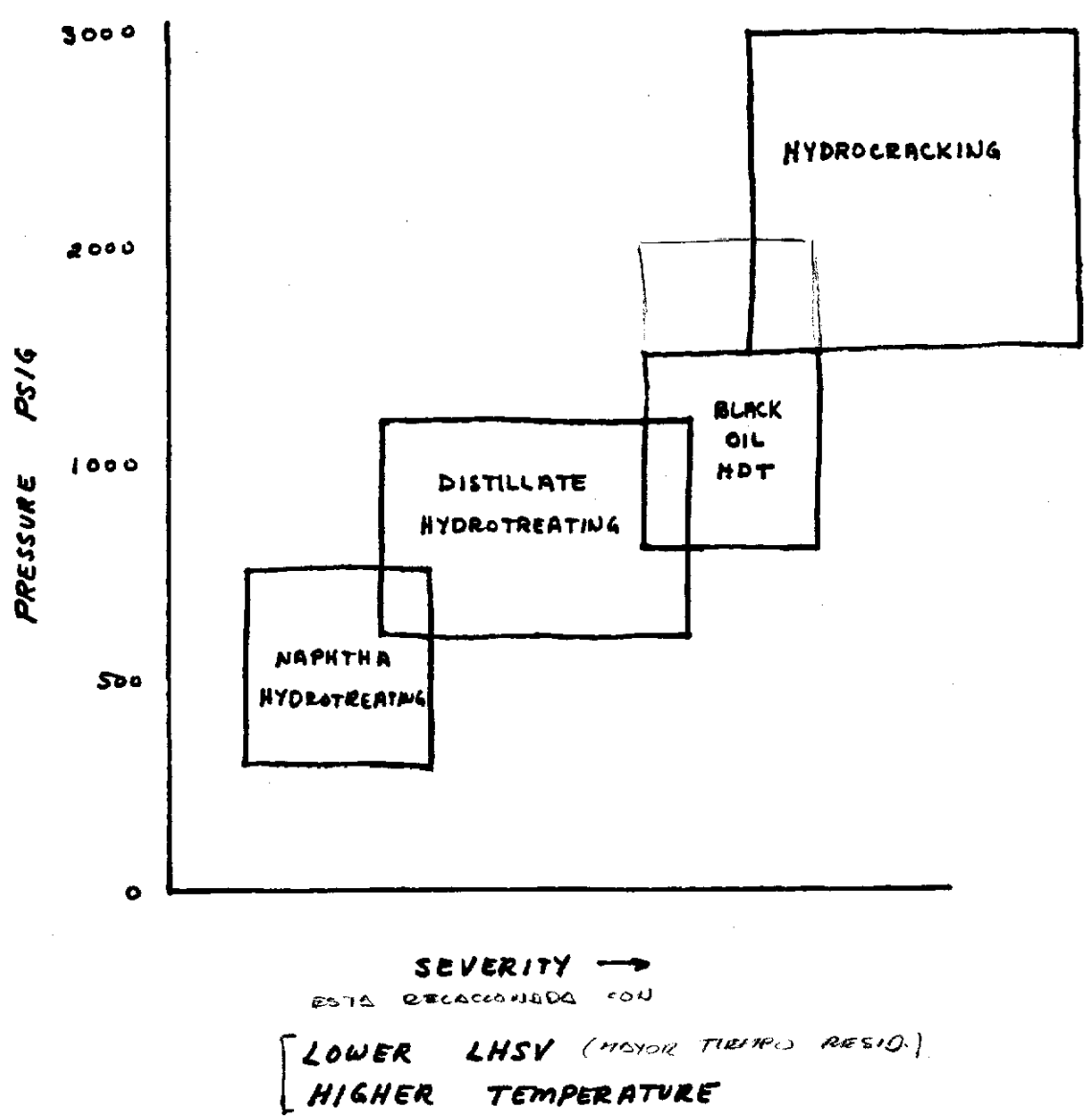
* CATALYSTS

COMPONENTES ACTIVOS

CO/H₂

N₂/H₂

CHEMISTRY OF HYDROPROCESSING



HYDROTREATING PROCESS VARIABLES

NORMAL RANGES

PRESSURE

CARGAS
LIVIANAS 17.6
CARGAS
PESADAS 210.9 kg/cm² G
250 - 3000 PSIG

TEMPERATURE

205°C 425°C
400 - 800 DEG F

LHSV

0.5 - 10

HYDROGEN RATE

250 - 10000 SCFB

HYDROTREATING PRODUCT YIELDS

REACTOR YIELD INFLUENCED BY

- : FEED PROPERTIES
- LEVEL OF DESULFURIZATION
- PROCESS CONDITIONS
- DESIRED PRODUCT PROPERTIES
- CATALYST TYPE

$$\text{WEIGHT YIELD} = \underbrace{\text{FEED}}_{W_{\text{FEED}}} + \underbrace{\text{HYDROGEN CONSUMED}}_{W_{\text{H}_2}}$$

DISTRIBUTION DE RENDIMIENTOS

NH₃ | INCLUYE

H₂S |

C1 - C5

C6 PLUS

MEASUREABLE VOLUMETRIC YIELD INCREASE

POR VARIACION DE DENSIDAD
DEBIDO A REARREGLOS MOLECULARES

HYDROPROCESSING UNIT DESIGN

*

MARKETING

COMMERCIAL TERMS

AGREEMENTS

BASIS

*

PROCESS DEVELOPMENT

BASIC FLOW SCHEME

CATALYSTS

PROCESS CONDITIONS

*

ENGINEERING

PROCESS

MATERIAL BALANCE

HEAT BALANCE

DISTILLATION

PROCESS FLOW AND INSTRUMENTATION

PROJECT

HYDRAULICS

METALLURGY

EQUIPMENT SPECIFICATIONS

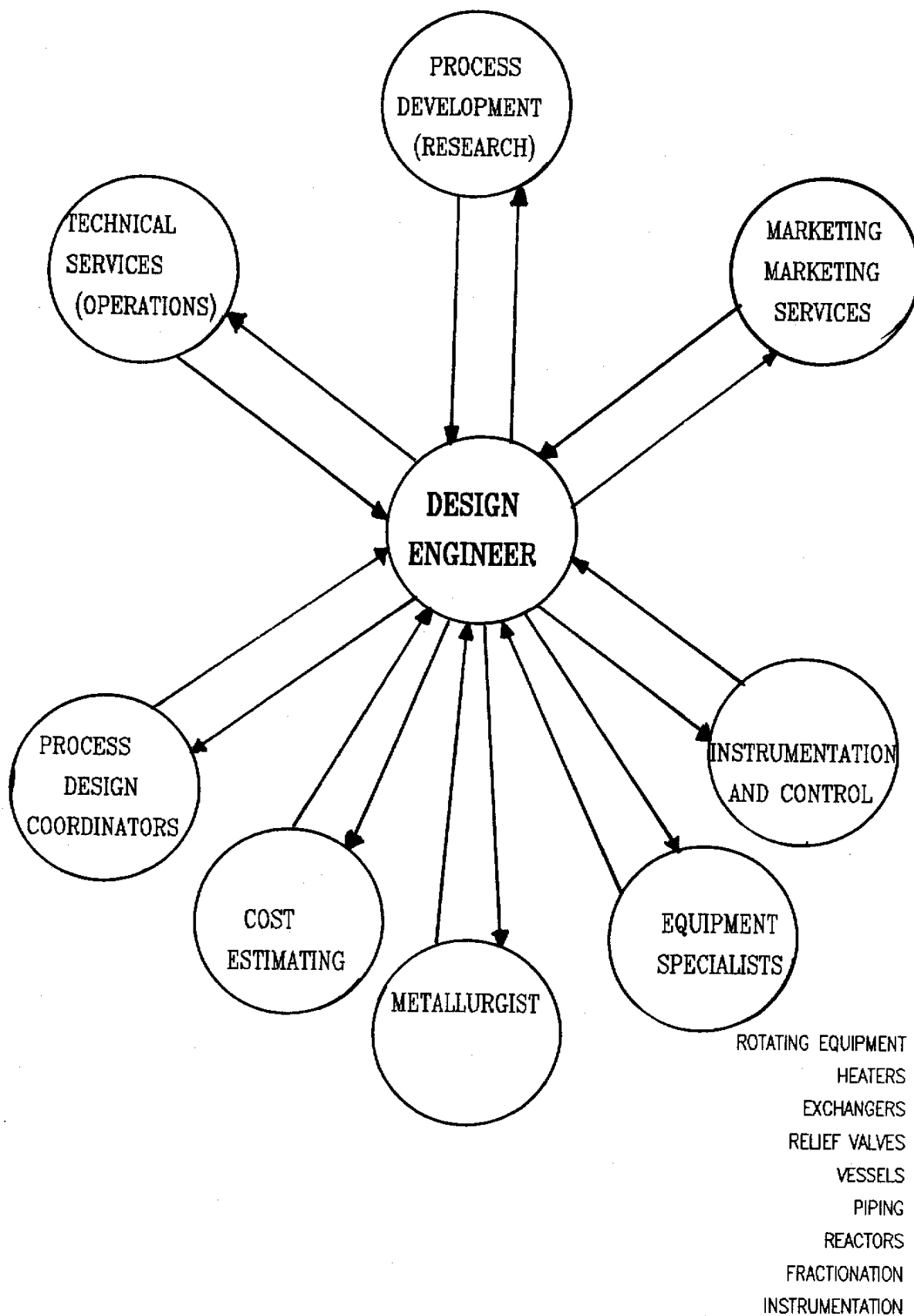
PIPING AND INSTRUMENT DIAGRAM

*

DETAILED ENGINEERING, PROCUREMENT, CONSTRUCTION

*

STARTUP AND OPERATION



PROCESS SIMULATION

*

MATHEMATICAL MODELS

LINEAR

NON - LINEAR

*

OPERATIONAL OPERATIONAL SIMULATION

GEN. ~~GENERAL~~ STEADY - STATE (SINGLE POINT)

DYNAMIC

MULTIVARIABLE

TIME DEPENDENT

*

ENGINEERING DESIGN SIMULATION

SE 803A BU

COMBINATIONS OF ABOVE MATH MOD
OPERAC. SIMULATION

FLEXIBILITY CONSIDERATIONS

OPERABILITY CONSIDERATIONS

SPECIFICATION OF EQUIPMENT

FEED AND PRODUCTS

AL DISEÑAR HIDROPROCESADORAS
ES IMPORTANTE
DEFINIR PAPA CARGA Y PRODUCTOS;

*

CHARACTERIZATION OF FEED OIL

ANALYSIS FUNDAMENTAL DE FEED STOCKS.

CORRELATIONS

THERMODYNAMIC PROPERTIES (PAPA COMP POROS Y HECK
DATOS EQUILIBRIO
Y CORRELACIONES

EXTENSIVE

INTENSIVE

*

DEFINITION OF CONTAMINANT LEVELS

FEED / PRODUCT

FEEDSTOCK PROCESSIBILITY

PROCESSING CONDITIONS

PROCESSING SCHEME

TIPO DE CONTAMINANTES
Y ESPECIFICACION DE
PRODUCTOS

*

CHARACTERIZATION OF PRODUCT OIL

PRODUCT SPECIFICATIONS (DADA POR ESTANDARIS)

PILOT PLANT / COMMERCIAL DATA

CORRELATIONS

*

PROCESSING HISTORY OF MATERIAL ANTES DE

LLGAR AL PROCESO DE HIDROTRATAMIENTO.

- TIPO DE TRATAMIENTO
- ALMACENAMIENTO

FEED STOCK DEFINITION

DATOS NECESARIOS DE LA CARGA

* CRUDE SOURCE OR DERIVATION OF FEEDS

* COMPOSITION

$\left. \begin{array}{l} C_1 \\ C_7 \end{array} \right\} \text{POR HRCG (CROMATOGRAFIA DE ALTA RESOLUCION)}$

* API GRAVITY (SPECIFIC GRAVITY)

* DISTILLATION

ASTM 52 DEBE TRANS A TBP
 TBP

* CONTAMINANT LEVELS

SULFUR

NITROGEN

BROMINE NUMBER (OLEFIN CONTENT)

CONRADSON CARBON

C7 INSOLUBLES (ASPHALTENES)

METALS

HALOGENS

NON - METALS

* MISC PROPERTIES

VISCOSITY

POUR POINT

COLOR

ACIDITY

ANILINE POINT

CETANE INDEX

FEED STOCK DEFINITIONS

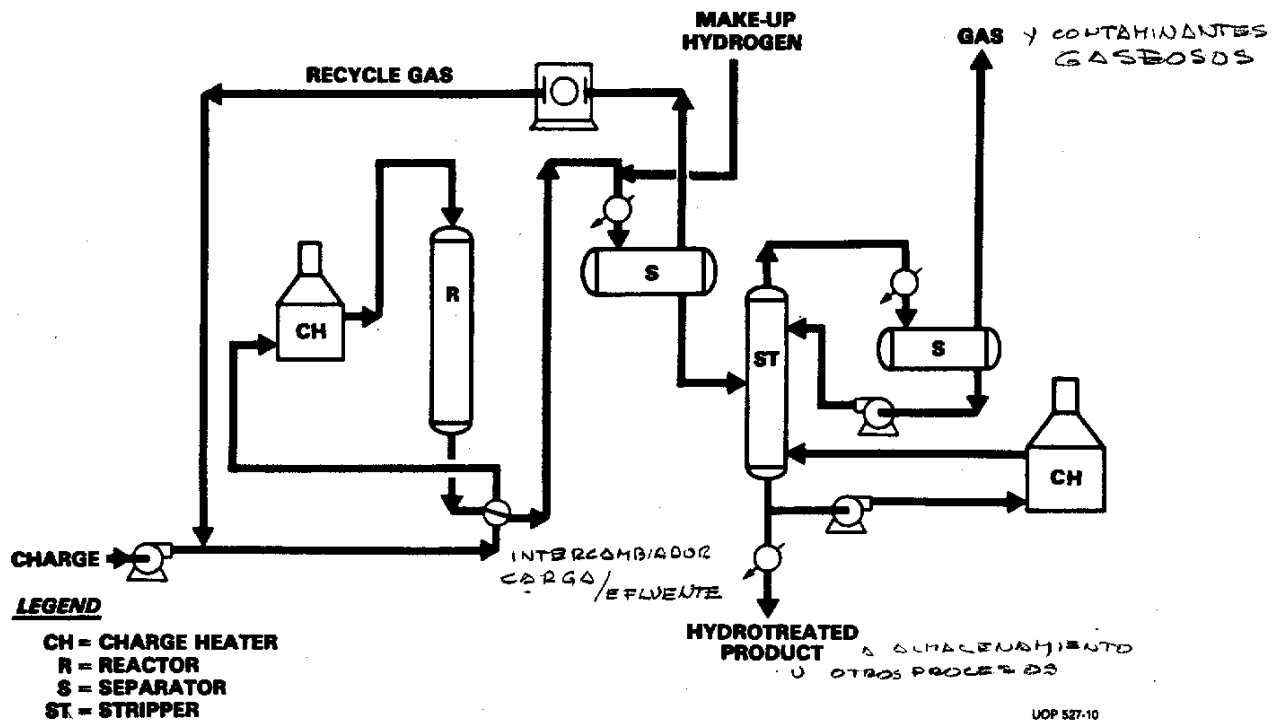
CONTAMINANT SOURCES

- * OIL FIELD PRODUCTION (SECONDARY / TERTIARY RECOVERY)
 - WATER FLOODING
 - SOLVENT TREATING

- * TRANSPORT AND STORAGE
 - SEA WATER CONTAMINATION
 - OXYGEN EXPOSURE
 - EXCESSIVE USE OF INHIBITOR

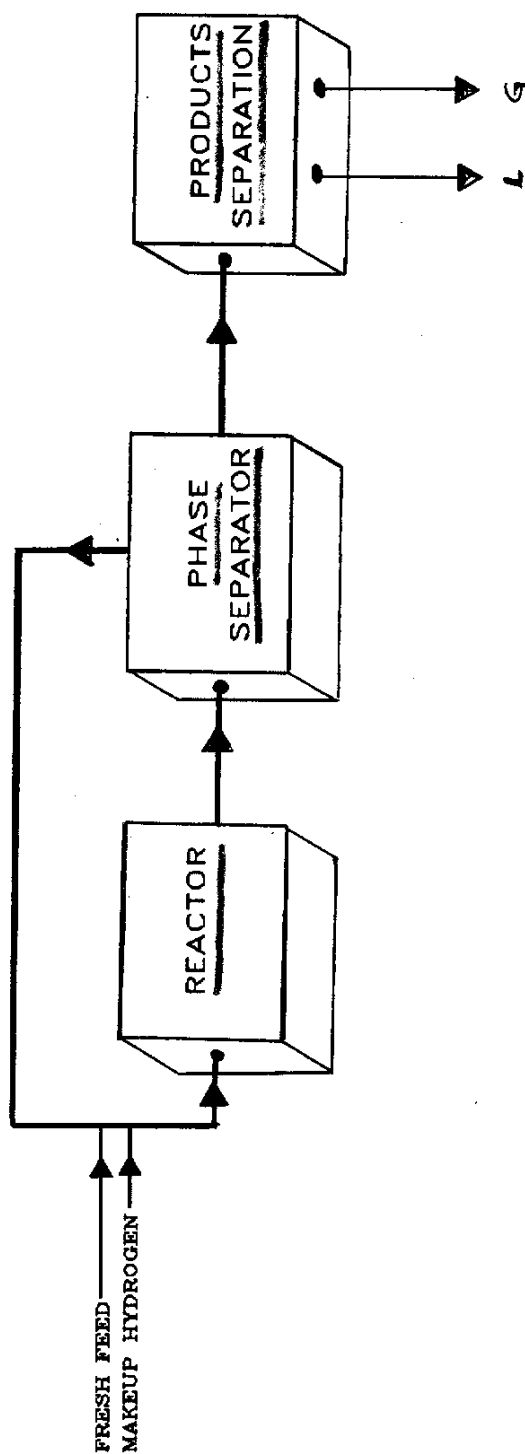
- * UPSTREAM OPERATIONS
 - INADEQUATE DESALTING
 - ENTRAINMENT FROM CRUDE / VACUUM OPERATIONS
 - ASPHALTENES
 - METALS
 - EXCESSIVE USE OF INHIBITOR
 - RERUN OF CONTAMINATED SLOPS

FIGURE I-1
NAPHTHA HYDROTREATING PROCESS



TIPICO ESQUEMA DE HIDROTREATAMIENTO

HYDROPROCESSING
(THE BASICS)
BLOQUES BASICOS DEL PROCESO



HYDROTREATING

PROCESS DESIGN - MATERIAL BALANCE

PARA ESTO SE NECESITA:

CHARGE STOCKS - CHARACTERIZATION

RAW OIL

MAKEUP HYDROGEN

REACTOR YIELD DISTRIBUTION

PRODUCTS - CHARACTERIZATION

PROCESSING CONDITIONS

MATERIAL BALANCE ENVELOPE

DEFINICION DEL SISTEMA

HYDROTREATING

NAPHTHA YIELD ESTIMATE

EJEMPLO

CHARGE STOCK CARACTERISTICAS.

SOURCE : STRAIGHT RUN LIGHT ARABIAN NAPHTHA

API	61.0	
SP GR	0.7350	
WT PCT S	0.032	
TN, PPM	1.0	NITROGENO TOTAL.
ASTM D-86 DEG-F		
		T °F
IBP	178	
10	220	
30	234	
50	250	
70	270	
90	300	
EP	330	

PRODUCT REQUIREMENTS OBJETIVO:

PLATFORMER FEED STOCK

SULFUR, WT PPM 0.5

TOTAL NITROGEN, PPM 0.5

HYDROTREATING

MAKEUP HYDROGEN COMPOSITION, MOL PCT

HYDROGEN	82.7
METHANE	6.3
ETHANE	4.7
PROPANE	3.1
BUTANES	1.8
PENTANES	1.4

HYDROTREATING NAPHTHA YIELD ESTIMATE

YIELD DISTRIBUTION

	WT PCT	VOL PCT
RAW OIL	100	100 LIQUID.
HYDROGEN	<u>0.05</u>	
	<u>100.05</u>	
NH3	TR	
H2S	0.03	
C1	TR	
C2	0.01	
C3	0.01	0.01
C4	0.02	0.02
C5	0.02	0.02
C6+	99.96	100.07
TOTAL	<u>100.05</u>	100.12 LIQUID
		CAMBIO DEBIDO A MODIFICACION ESTRUCTURAL

PRODUCT SPECIFIC GRAVITY

C4 SP GR 0.5787 ESTIMADO
C5 SP GR 0.6290 "
C6+ API 61.2 DE LA CARACT DE C6+

HYDROTREATING NAPHTHA YIELD ESTIMATE

PROCESS CONDITIONS

PRESSURE AT SEPARATOR, PSIG	350
MINIMUM REACTOR INLET TEMPERATURE, DEG-F	600
MAXIMUM REACTOR OUTLET TEMPERATURE, DEG-F	700
FRESH FEED RATE, BPSD	10000
LSHV ON FRESH FEED	8
H2 TO REACTOR INLET ON FF, SCFB	400
MAXIMUM ALLOWABLE RX DELTA-T, DEG-F	100
* ESPECIFIC MINIMUM H2 IN RECYCLE GAS, MOL PCT	70
HEAT OF REACTION ON FF, BTU/LB	6.5
WASH WATER, VOL PCT ON FF	3
ESTIMATED SERVICE, 1ST CYCLE	1 YEAR
ULTIMATE	5 YEARS

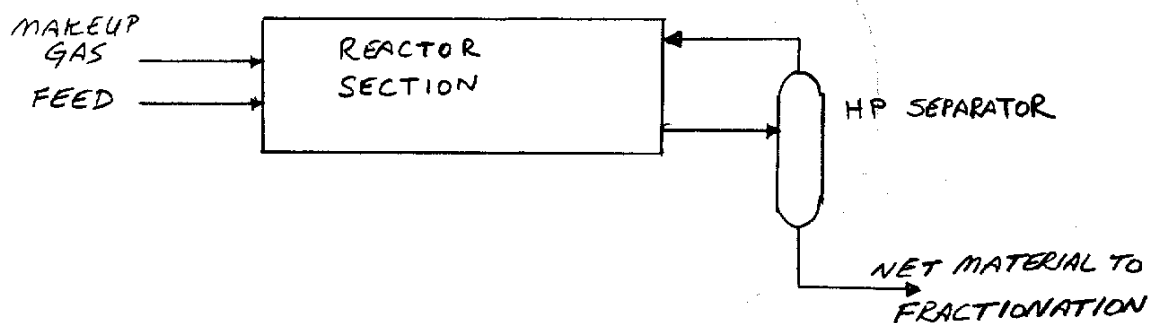
A TRAVES
DEL RA.

PARA POSICION
REACTOR 12 12

* PARA MINIMIZAR CORROSION DE LA PARTE PRIN DE LA
PLANTA.

HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

SYSTEM



TYPICAL FLOW SCHEME

ALL MATERIAL IN HP SEPARATOR LIQUID

MAKEUP HYDROGEN REQUIREMENT IS MINIMUM

HP SEPARATOR AT BUBBLE POINT

HP SEPARATOR EQUILIBRIUM INDEPENDENT OF RECYCLE GAS RATE

HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

VAPOR - LIQUID EQUILIBRIUM

- f_i = mols of component i in feed
- F = Total mols of feed
- v_i = mols of component i in vapor
- V = Total mols of vapor
- l_i = mols of component i in liquid
- L = Total mols of liquid
- z_i = mol fraction of component i in feed (f_i / F)
- y_i = mol fraction of component i in vapor (v_i / V)
- x_i = mol fraction of component i in liquid (l_i / L)
- K_i = thermodynamic equilibrium constant (y_i / x_i)

For dew point vapor $\sum y_i / K_i = 1.0 \quad \sum z_i / K_i = V$

For bubble point liquid $\sum K_i x_i = 1.0 \quad \sum K_i z_i = L$

For partial vaporization $\frac{V}{V + L} = \sum \frac{K_i z_i}{K_i + L/V}$

HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

FEED NAPHTHA CHARACTERIZATION

$$VABP = \frac{10\% + 30\% + 50\% + 70\% + 90\%}{5}$$

$$= (220 + 234 + 250 + 270 + 300) / 5$$

$$= 254.8$$

$$ENGLER \ SLOPE = (90\% - 10\%) / 80$$

$$= (300 - 220) / 80 = 1.0$$

$$API = 61 \Rightarrow 6.1286 \text{ LB/GAL}$$

$$MABP = 250.2 \Rightarrow MW = 112.6$$

$$CABP = 253.3 \Rightarrow VOP \ K = \underline{12.15}$$

DE LOS
CARTAS
DE CORRECTE
RIZACION
DE HC.

PRODUCT C6+ CHARACTERIZATION

ASSUME ASTM D-86 OF PRODUCT $\approx$ FEED

$$API = 61.2 \Rightarrow 6.122 \text{ LB/GAL}$$

$$VABP = 254.8 \quad SLOPE = 1.0$$

$$MABP = 250.2 \Rightarrow MW = \underline{112.7}$$

$$CABP = 253.3 \Rightarrow VOP \ K = \underline{12.16}$$

DE LOS
CARTAS
DE CORRECTE
RIZACION

3:

HYDROTREATING
PROCESS DESIGN - MATERIAL BALANCE

$$\text{IN} = \text{OUT} + \text{ACCUMULATION}$$

STEADY STATE OPERATION

$$\text{ACCUMULATION} = 0$$

MASS CONSERVED ACROSS REACTOR



HYDROTREATING
PROCESS DESIGN - MATERIAL BALANCE

1) FEED MASS RATES

$$\begin{aligned}\text{RAW OIL RATE} &= 10,000 \text{ BPSD} \\ &= (10000)(42)(6.1286)/24 \\ &= 107250 \text{ LBS/HR}\end{aligned}$$

$$\begin{aligned}\text{HYDROGEN CONSUMPTION} &= (107250)(.0005) \\ &= 53.625 \text{ LBS/HR}\end{aligned}$$

$$\text{TOTAL} = 107304 \text{ LBS/HR}$$

MASS CONSERVED ACROSS REACTOR

$$\text{IN} = \text{OUT} = 107304 \text{ LBS/HR}$$

HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

2) REACTOR YIELDS

	WT% x CONVERSION	LBS/HR ÷ MW	MOL/HR
H ₂ S	0.03 x 107304/100 = 32.19 ÷ 34.076 = 0.94		
H ₂	-0.05	-53.65	2.016 -26.61
C ₁	-	-	-
C ₂	0.01	10.73	30.07 0.36
C ₃	0.01	10.73	44.1 0.29
C ₄	0.02	21.46	58.124 0.37
C ₅	0.02	21.46	72.146 0.30
C ₆ ⁺	99.96	107261.08	112.7 951.78
TOTAL	100.00	107304	927.38

3) PRODUCT SEPARATOR EQUILIBRIUM DADA DETERMINAR CONSUMO DEL GAS DE REPOSICION MOL BALANCE

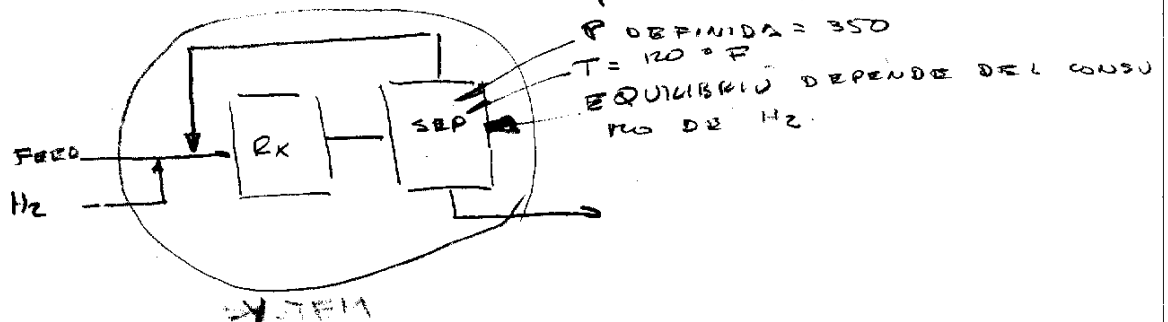
$$(\text{REACTOR YIELD}) + (\text{MAKEUP H}_2) = (\text{SEPARATOR LIQUID})$$

$$\text{BUBBLE POINT MATERIAL IN SEPARATOR } \sum x_i \cdot K_i = 1.0$$

SET

TEMPERATURE = 120 DEG-F (ECONOMICS) SELECCIONADA DE ACUERDO A CONVENIENCIA ECONOMICA

PRESSURE = 350 PSIG (PROCESS REQUIREMENT)



HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

ASSUME MAKEUP $H_2 = 40 \text{ mol/hr}$ FORDS TOWER BL EQUILIBRIUM

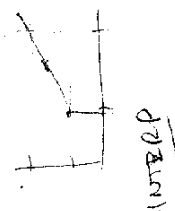
TRIAL 1

COMP	P-X YIELD	MAKEUP H_2	SEP LIQ	X_i	K_i	$X_i K_i$
H_2S	0.94	0	0.94	9.72×10^{-4}	1.37	0.0013
H_2	-26.61	33.07	6.46	6.68×10^{-3}	50	0.3339
C1	-	2.49	2.49	2.57×10^{-3}	7.81	0.0201
C2	0.36	1.89	2.25	2.33×10^{-3}	1.96	0.0046
C3	0.24	1.25	1.49	1.54×10^{-3}	0.65	0.0010
C4	0.37	0.73	1.10	1.14×10^{-3}	0.247	0.0003
C5	0.30	0.57	0.87	8.99×10^{-4}	0.079	0.0001
C6+	951.78	0	951.78	0.9839	0.004	0.0039
TOTAL	927.38	40.0	967.38			<u>0.3652</u> < 1.0.

HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

		ASSUME MAKEUP $H_2 = 60 \text{ MOL/Hr}$			
		<u>TRIAL 2</u>			
COMP	LX YIELD	MAKEUP H_2	SEP LIQ	X_i	K_i
H_2S	0.94	0	0.94	9.52×10^{-4}	1.37
H_2	26.61	49.60	22.99	2.33×10^{-2}	50.0
C1	-	3.74	3.74	3.79×10^{-3}	7.81
C2	0.36	2.84	3.20	3.24×10^{-3}	1.96
C3	0.24	1.88	2.12	2.15×10^{-3}	0.65
C4	0.37	1.09	1.46	1.48×10^{-3}	0.247
C5	0.30	0.85	1.15	1.16×10^{-3}	0.079
C6+	951.78	0	951.78	0.9639	0.004
TOTAL		60.0	987.38	1.2034	
				=	
				> 1.0	

40 - 0.36
60 - 1.02
60 - 1.20



HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

ASSUME MAKEUP $H_2 = 55 \text{ mol/hr}$

TRIAL 3

COMP	RX YIELD	MAKEUP H_2	SEP LIQ	X_i	K_i	$X_i K_i$	y_i PCT
H_2S	0.94	0	0.94	9.57×10^{-4}	1.37	0.0013	0.13
H_2	-26.61	45.47	18.86	1.92×10^{-2}	50.0	0.9599	95.99
C1	-	3.43	3.43	3.49×10^{-3}	7.81	0.0273	2.73
C2	0.36	2.60	2.96	3.01×10^{-3}	1.96	0.0059	0.59
C3	0.24	1.72	1.96	2.0×10^{-3}	0.65	0.0013	0.13
C4	0.37	1.00	1.37	1.39×10^{-3}	0.247	0.0003	0.03
C5	0.30	0.78	1.08	1.10×10^{-3}	0.079	0.0001	0.01
C6+	951.78	0	951.78	0.9685	0.004	0.0039	0.39
TOTAL	927.38	55.00	982.38			1.0	100

HYDROTREATING PROCESS DESIGN - MATERIAL BALANCE

COMP	RX YIELD	MAKEUP H ₂	SEP LIQ.	X _i	K _i	X _i K _i	Y _i PCT
H ₂ S	0.94	0	0.94	9.57×10^{-4}	1.37	0.0013	0.13
H ₂	-26.61	45.47	18.86	1.92×10^{-4}	50.0	0.9599	95.99
C1	-	3.43	3.43	3.49×10^{-3}	7.81	0.0273	2.73
C2	0.36	2.60	2.96	3.01×10^{-3}	1.96	0.0059	0.59
C3	0.24	1.72	1.96	2.0×10^{-3}	0.65	0.0013	0.13
C4	0.37	1.00	1.37	1.39×10^{-3}	0.247	0.0003	0.03
C5	0.30	0.78	1.08	1.10×10^{-3}	0.079	0.0001	0.01
C6+	951.78	0	951.78	0.96885	0.004	0.0039	0.39
TOTAL	927.38	55.00	982.38			1.0	100

GAS BECULUO

4

HYDROTREATING

PROCESS DESIGN - MATERIAL BALANCE

RECYCLE GAS RATE

$$\begin{aligned}\underline{H_2 \text{ TO REACTOR}} &= 400 \text{ SCFB} \\ &= (400)(10000) / (24)(379) \\ &= 439.75 \text{ MOL/HR}\end{aligned}$$

$$\begin{aligned}\underline{\text{RECYCLE GAS TO RX}} &= (439.75) / (0.9599) \\ &= 458.12 \text{ MOL/HR} \\ &= 460 \text{ MOL/HR}\end{aligned}$$

RECYCLE GAS COMPONENT RATE

COMPONENT	COMPOSITION MOL FRACTION	x FACTOR	<u>GAS DE RECYCLE</u> <u>MOL/HR</u>
H ₂ S	0.0013	x 460	0.60
H ₂	0.9599		441.55
C ₁	0.0273		12.56
C ₂	0.0059		2.71
C ₃	0.0013		0.60
C ₄	0.0003		0.14
C ₅	0.0001		0.05
C ₆ ⁺	0.0039		1.79
TOTAL	1.0		460.

HYDROTREATING

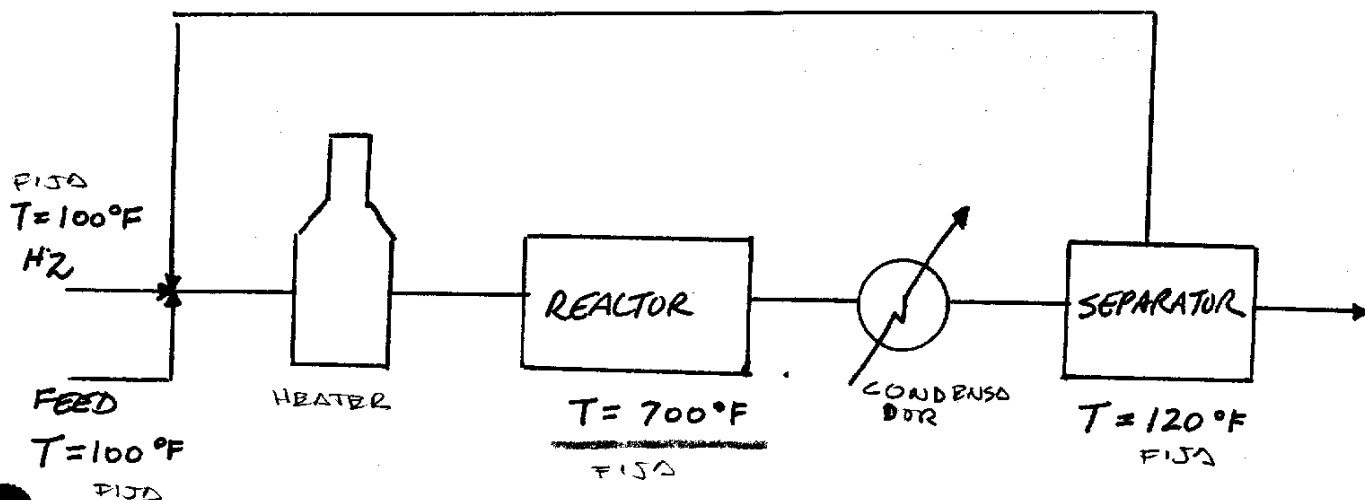
PROCESS DESIGN - MATERIAL BALANCE

	① FRESH FEED	② RECYCLE GAS	③ CF TO RX	④ RX YLD	⑤ RX EFFL	⑥ MAKEUP H ₂	⑦ MTL TO SEP	⑧ SEP LIQ	⑨ SEP VAPOR
COMP									
H ₂ S		0.6	0.6	0.94	1.54	-	1.54	0.94	0.60
H ₂	441.55		441.55	-26.61	414.94	45.47	460.41	18.86	441.55
C1	12.56		12.56	-	12.56	3.43	15.99	3.43	12.56
C2	2.71		2.71	0.36	3.07	2.60	5.67	2.96	2.71
C3	0.60		0.60	0.24	0.84	1.72	2.56	1.96	0.60
C4	0.14		0.14	0.37	0.51	1.00	1.51	1.37	0.14
C5	0.05		0.05	0.30	0.35	0.78	1.13	1.08	0.05
RAWOIL	951.64	-	951.64	-951.64	0	0	0	0	-
CGT	1.79		1.79	951.78	953.57	0	953.57	951.78	1.79
TOTAL	951.64	460	1411.64		1387.38	55	1442.38	982.38	460
LB/HAL	107250	1433	108683		108683	415	109278	107845	1433
BPSP	10000	-						10094	
MW	112.7	3.12	76.72		76.34	7.54	75.76	109.78	3.12
MMSLED		4.184				1.00			4.184

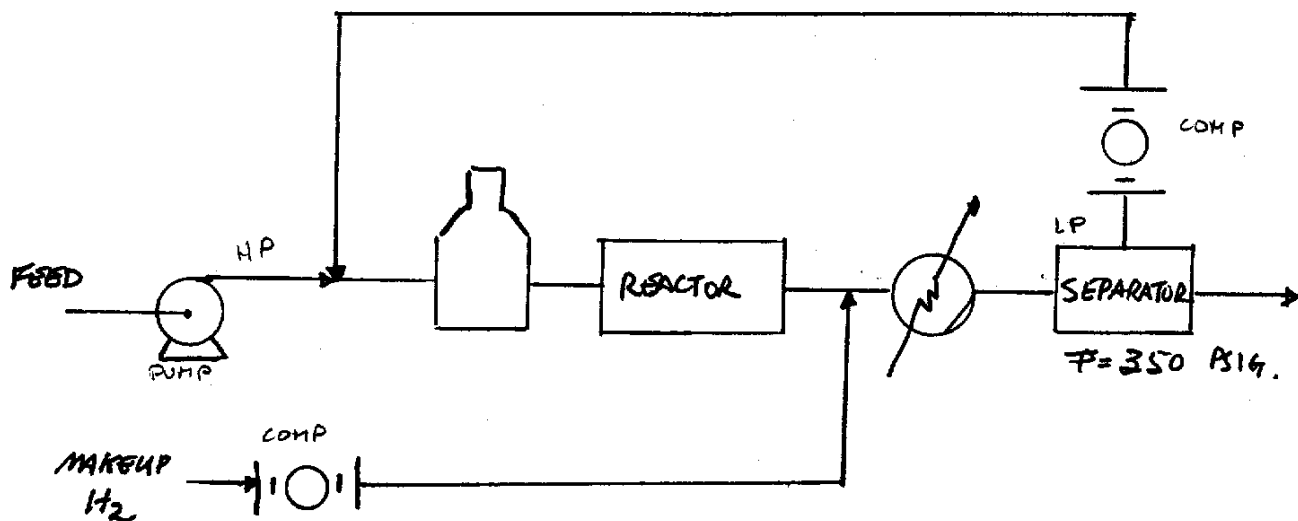
HYDROTREATING HEAT BALANCE

REACTOR - THERMAL REQUIREMENTS

2 EQUIPOS

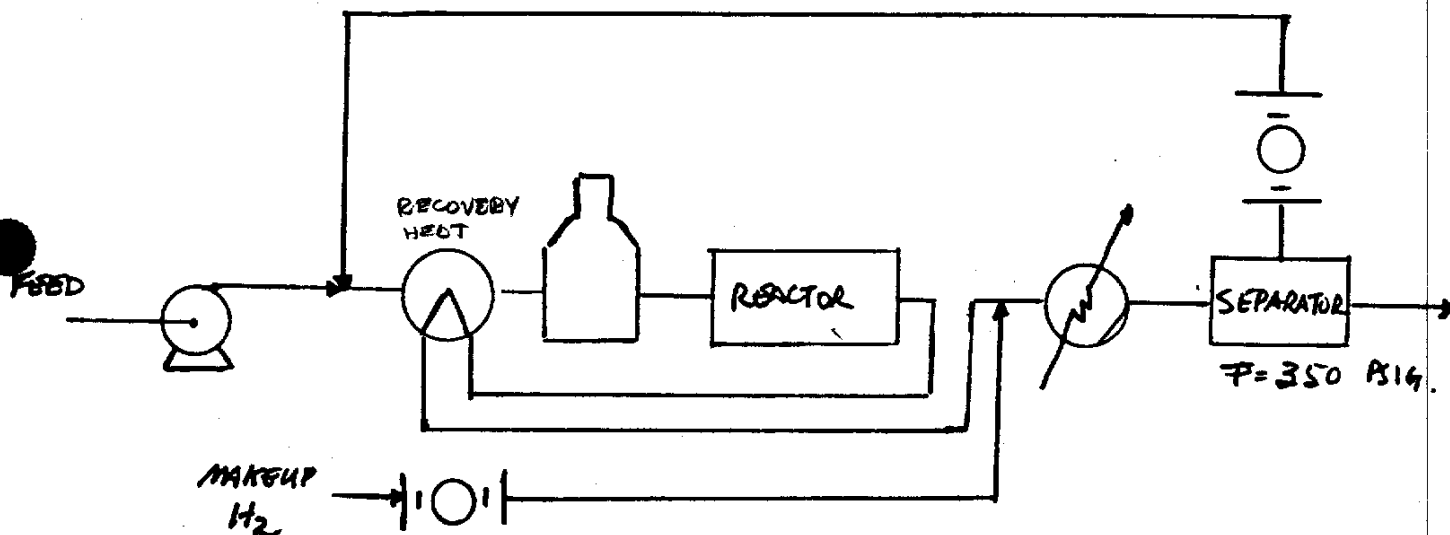


HYDRAULICS REQUIREMENTS



HYDROTREATING HEAT BALANCE

EFFICIENCY - WASTE HEAT RECOVERY



SI INCREMENTA TEMP RX

SE INCREMENTA RECUPERACION DE CALOR

INGRESO DE CALOR [VAPOR
ACETOS (LOW THERM) PETROQUIMICAS
DIRECT FIRE HEATERS (REFINERIAS)

50

HYDROTREATING HEAT BALANCE

PROCESS REQUIREMENTS

REACTION TEMPERATURE

SOR - MINIMUM RX INLET

EOR - MAXIMUM RX OUTLET

UOP DESIGNS FOR ^{END OF REACTOR} EOR AND
EVALUATES OTHER OPERATIONS

HEAT OF REACTION

HYDROTREATING REACTIONS EXOTHERMIC

UOP EXPRESSES AS BTU/LB FEED

HYDROTREATING HEAT BALANCE

HYDRAULICS

NEED TO ASSIGN PRELIMINARY ESTIMATES
OF EQUIPMENT, VALVE, LINE PRESSURE
DROPS FOR PROCESS DESIGN PURPOSES

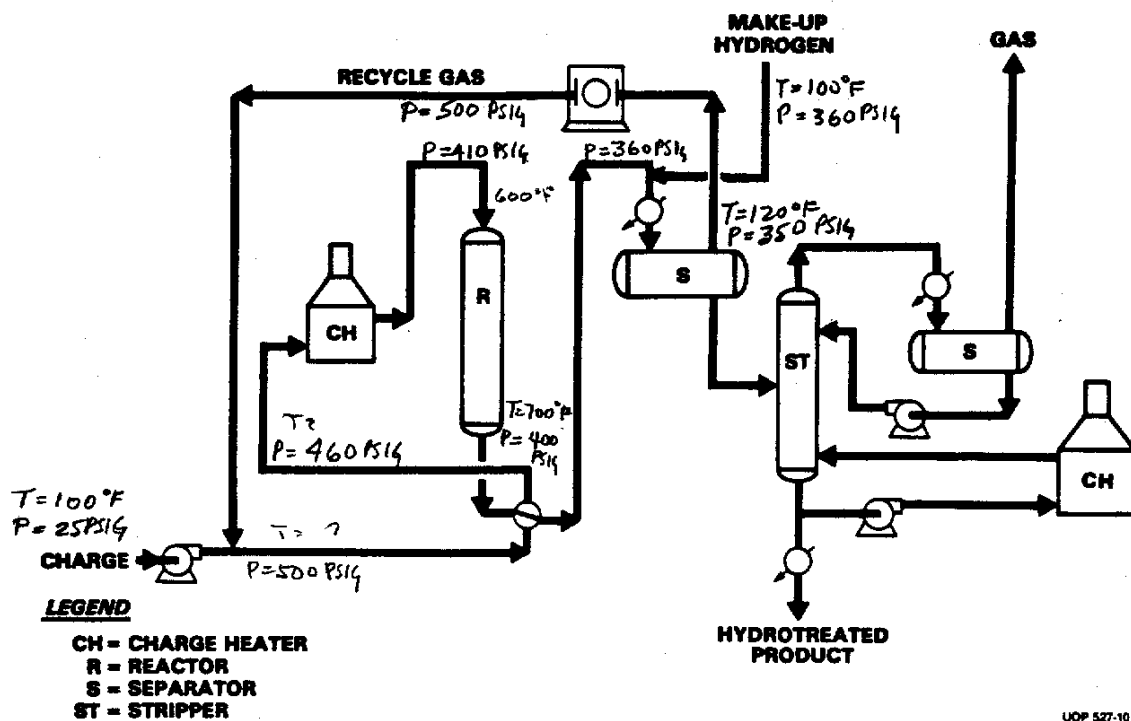
AT OPERATING PRESSURES IN NAPHTHA
HYDROTREATING, PRESSURE DIFFERENTIALS
IN REACTOR CIRCUIT HAVE MINIMAL
EFFECT ON EQUILIBRIUM 'K' AND
ENTHALPY FUNCTION.

HYDROTREATING HEAT BALANCE

HYDRAULICS TYPICAL FOR NHIT.

	<u>ΔP</u>	PRESSURE <u>PSIG.</u>
SEPARATOR		<u>350</u> CONTROL POINT.
CONDENSER	<u>10</u>	360
EXCHANGE (TUBE)	<u>40</u>	400
REACTOR	<u>10</u>	410
HEATER	<u>50</u>	460
EXCHANGE (SHELL)	<u>40</u>	500

FIGURE I-1
NAPHTHA HYDROTREATING PROCESS



UOP 527-10

HYDROTREATING. HEAT BALANCE.

RECYCLE GAS COMPRESSOR.

ADDS ENERGY TO SYSTEM.

TEMPERATURE RISE ON COMPRESSION.

TYPE OF COMPRESSOR SELECTION

SYSTEM REQUIREMENTS

CAPACITY REQUIREMENTS

SUCTION CONDITIONS $T = 120^{\circ}\text{F}$ (580°R)
 $P = 350 + 14.7 = 364.7 \text{ PSIA}$
 $Z \approx 1.0$
 $C_p/C_v \approx 1.27.$

1) CALCULATE ACFM

$$V = \frac{ZNR T}{P} = \frac{(1.0)(460)(10.73)(580)}{364.7}$$

$$= 7849.64 \text{ FT}^3/\text{HR}$$

$$\text{ACFM} = \underline{130.8 \text{ FT}^3/\text{MIN}}$$

HYDROTREATING HEAT BALANCE.

RECYCLE COMPRESSOR.

SELECTED MACHINE - RECIPROCATING

2) ESTIMATE DISCHARGE TEMPERATURE

$$P = 500 + 14.7 = 514.7.$$

$$\text{COMPRESSION RATIO} = \frac{514.7}{364.7} = 1.4113$$

$$K \approx C_p/C_v = 1.37.$$

DISCHARGE TEMPERATURE

$$\begin{aligned} T_2 &\approx T_1 (P_2/P_1)^{K-1/K} \\ &= (580)(1.4113)^{(1.37-1)/1.37} \\ &= 636.6 \text{ } ^\circ\text{R} \\ &= 176.6 \text{ } ^\circ\text{F} \\ &\approx \underline{177 \text{ } ^\circ\text{F}.} \end{aligned}$$

HYDRO TREATING. HEAT BALANCE

REACTOR CONDITIONS - OUTLET FOR
PHASE CONDITION
ENTHALPY OF MIXTURE

$$T = 700^{\circ}\text{F} -$$

$$P = 400 \text{ PSIG}$$

	RX EFF	y_i	POINT 700°F K_i	y_i/K_i
H ₂ S	1.54	1.11×10^{-3}	564	2.08×10^{-4}
H ₂	414.94	0.2991	1154	0.0272
C ₁	12.56	9.053×10^{-3}	65	1.393×10^{-3}
C ₂	3.07	2.213×10^{-3}	4.6	4.81×10^{-4}
C ₃	0.84	6.055×10^{-4}	4.3	1.408×10^{-4}
C ₄	0.51	3.676×10^{-4}	3.5	1.05×10^{-4}
C ₅	0.35	2.553×10^{-4}	3.0	8.51×10^{-5}
C ₆ ⁺	953.57	0.6873	2.4	0.2864
	<u>1387.38</u>			

$$0.316 < 1.0$$

MATERIAL SALIENDO DEL REACTOR
PODA TODO EN FASE GASOSA.
ALL VAPOR.

HYDROTREATING HEAT BALANCE

REACTOR	EFFLUENT	ENTHALPY
	$\frac{\text{MW}}{\text{EFF}}$ LB/HR	$\frac{\text{BTU/LB}}{700^{\circ}\text{F}}$ BTU/HR.
H_2S	1.54 52.36	364 19074
H_2	414.90 830	2200 1826000
C_1	12.56 201	507 101907
C_2	3.07 92	506 46552
C_3	0.84 37	524 19388
C_4	0.51 30	516 15480
C_5	0.35 25	516 12900
C_6^+	95357 107467	496 53303632
		<u>55.35×10^6</u>

HEAT OF REACTION HEAT RELEASE.

$$\text{BTU/HR} = \frac{\Delta H_{\text{RX}} \text{ BTU/LB.}}{(107250)(6.5)} = 0.697 \times 10^6 \text{ BTU/HR.}$$

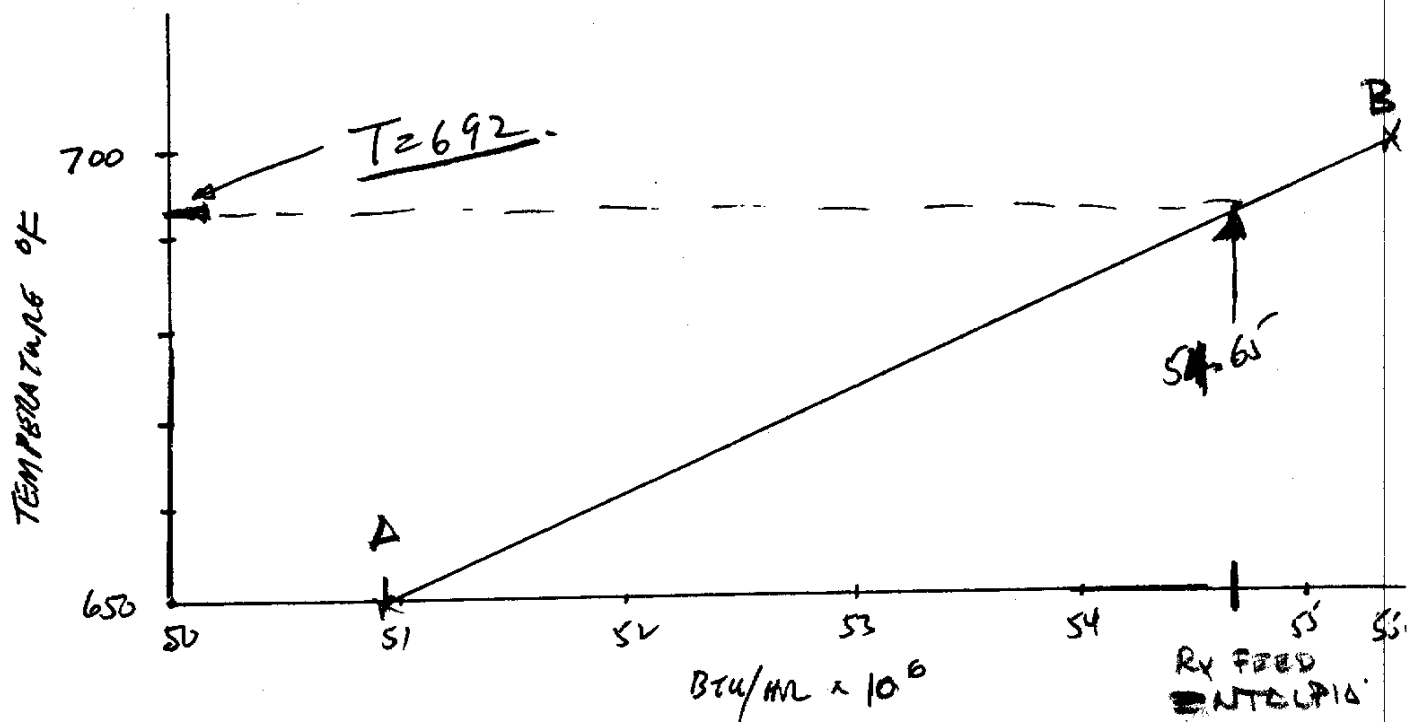
$$\begin{aligned} \text{REACTOR FEED ENTHALPY} &= 55.35 \times 10^6 - 0.697 \times 10^6 \\ &= \underline{54.65 \times 10^6 \text{ BTU/HR.}} \end{aligned}$$

BY TRIAL AND ERROR TEMPERATURE
OF FEED TO REACTOR CAN BE CALCULATED

HYDROTREATING HEAT BALANCE

REACTOR INLET TEMPERATURE

	FEED mol/hr	LB/hr	200°F ENTHALPY BTU/LB	BTU/hr	650°F BTU/LB	BTU/hr
H_2S	0.6	20.4	365	7446	350	7140
H_2	441.55	883.1	2200	1.942×10^6	2030	1.7927×10^6
C_1	12.56	201	507	101907	467	93867
C_2	2.71	81.3	506	41138	480	39024
C_3	0.60	26.4	524	13834	486	39512
C_4	0.14	6.16	516	3179	468	2883
C_5	0.05	3.6	516	1858	476	1714
C_6^+	1.79	201.7	496	100043	456	91975
FEED.	456.57	107250	496	53.196×10^6	456	48.906×10^6
				55.407×10^6		50.975×10^6
				B		A



HYDROTREATING

Heat Balance

makeup hydrogen enthalpy at 100

mkuph2	f	MW	H	MMBtu/Hr
h2	45.47	2	133.8	0.012
c1	3.43	16	101.3	0.006
c2	2.60	30	164.4	0.013
c3	1.72	44	179.3	0.014
c4	1.00	58	173.3	0.010
c5	<u>0.78</u>	72	171.8	<u>0.010</u>
total	55.00			0.064

fresh feed enthalpy at 100

ffeed	f	MW	h	MMBtu/Hr
c6+	951.64	112.7	19.7	<u>2.112821</u>

recycle gas enthalpy at 177

rcy gas	v	MW	H	MMBtu/Hr
h2s	0.60	34	221	0.005
h2	441.55	2	380	0.336
c1	12.56	16	138	0.028
c2	2.71	30	190	0.015
c3	0.60	44	204	0.005
c4	0.14	58	190	0.002
c5	0.05	72	190	0.001
c6+	<u>1.79</u>	112.7	136	<u>0.027</u>
total	460.00			0.418

HYDROTREATING Heat Balance

dew point calc of combined feed

cfeed	f	K500	v	K450	v	K460	v
h2s	0.60	6.50	0.09	7.60	0.08	7.60	0.08
h2	441.55	17.50	25.23	20.00	22.08	19.50	22.64
c1	12.56	11.30	1.11	12.20	1.03	12.10	1.04
c2	2.71	5.40	0.50	5.40	0.50	5.40	0.50
c3	0.60	2.90	0.21	3.60	0.17	3.65	0.16
c4	0.14	2.75	0.05	2.50	0.06	2.55	0.05
c5	0.05	1.82	0.03	1.52	0.03	1.60	0.03
<u>c6+</u>	<u>953.43</u>	<u>0.98</u>	<u>972.89</u>	<u>0.62</u>	<u>1537.79</u>	<u>0.69</u>	<u>1390.04</u>
total	1411.64		1000.11		1561.73		1414.56
y			0.7085		1.1063		<u>1.0021</u>

dew point calc of rx effluent

rxeff	f	K460	v	K450	v	K463	v
h2s	1.54	7.60	0.20	7.60	0.20	7.60	0.20
h2	414.94	19.50	21.28	20.00	20.75	19.25	21.56
c1	12.56	12.10	1.04	12.20	1.03	12.05	1.04
c2	3.07	5.40	0.57	5.40	0.57	5.40	0.57
c3	0.84	3.65	0.23	3.60	0.23	3.67	0.23
c4	0.51	2.55	0.20	2.50	0.20	2.56	0.20
c5	0.35	1.60	0.22	1.52	0.23	1.61	0.22
<u>c6+</u>	<u>953.57</u>	<u>0.69</u>	<u>1381.99</u>	<u>0.62</u>	<u>1538.02</u>	<u>0.71</u>	<u>1352.58</u>
total	1387.38		1405.72		1561.23		1376.60
y			1.0132		1.1253		<u>0.9922</u>

HYDROTREATING Heat Balance

cf vapor-liquid split at 300

cfeed	f	K300	f*K300	K+L/V	L/V=2	K+L/V	L/V=1.93
					v		v
h2s	0.60	4.70	2.82	6.70	0.42	6.63	0.43
h2	441.55	29.60	13069.88	31.60	413.60	31.53	414.52
c1	12.56	11.90	149.46	13.90	10.75	13.83	10.81
c2	2.71	4.45	12.06	6.45	1.87	6.38	1.89
c3	0.60	2.18	1.31	4.18	0.31	4.11	0.32
c4	0.14	1.39	0.19	3.39	0.06	3.32	0.06
c5	0.05	0.60	0.03	2.60	0.01	2.53	0.01
<u>c6+</u>	<u>953.43</u>	<u>0.12</u>	<u>109.64</u>	<u>2.12</u>	<u>51.84</u>	<u>2.05</u>	<u>53.62</u>
total	1411.64				478.87		481.65
L/V					1.948		<u>1.931</u>

cf vapor-liquid split at 400

cfeed	f	K400	f*K400	K+L/V	L/V=1.1	K+L/V	L/V=1.07
					v		v
h2s	0.60	7.00	4.20	8.10	0.52	8.07	0.52
h2	441.55	22.80	10067.34	23.90	421.23	23.87	421.76
c1	12.56	13.00	163.28	14.10	11.58	14.07	11.60
c2	2.71	5.20	14.09	6.30	2.24	6.27	2.25
c3	0.60	3.20	1.92	4.30	0.45	4.27	0.45
c4	0.14	2.18	0.31	3.28	0.09	3.25	0.09
c5	0.05	1.22	0.06	2.32	0.03	2.29	0.03
<u>c6+</u>	<u>953.43</u>	<u>0.37</u>	<u>352.77</u>	<u>1.47</u>	<u>239.98</u>	<u>1.44</u>	<u>244.98</u>
total	1411.64				676.11		681.68
L/V					1.088		<u>1.071</u>

HYDROTREATING Heat Balance

combined feed enthalpy at 300

cfeed	v	MW	H	MMBtu/Hr	I	MW	h	MMBtu/Hr
h2s	0.43	34	253.0	0.004	0.17	34	183.9	0.001
h2	414.52	2	815.6	0.676	27.03	2	183.9	0.010
c1	10.81	16	213.1	0.037	1.75	16	183.9	0.005
c2	1.89	30	250.9	0.014	0.82	30	183.9	0.005
c3	0.32	44	256.4	0.004	0.28	44	183.9	0.002
c4	0.06	58	238.2	0.001	0.08	58	183.9	0.001
c5	0.01	72	245.5	0.000	0.04	72	153.7	0.000
<u>c6+</u>	<u>53.62</u>	112.7	258.8	<u>1.564</u>	<u>899.81</u>	112.7	135.0	<u>13.690</u>
total	481.65			2.299	929.99			13.715

total enthalpy

16.014

combined feed enthalpy at 400

cfeed	v	MW	H	MMBtu/Hr	I	MW	h	MMBtu/Hr
h2s	0.52	34	280.9	0.005	0.08	34	204.7	0.001
h2	421.76	2	1161.1	0.979	19.79	2	204.7	0.008
c1	11.60	16	279.9	0.052	0.96	16	204.7	0.003
c2	2.25	30	311.0	0.021	0.46	30	204.7	0.003
c3	0.45	44	317.8	0.006	0.15	44	204.7	0.001
c4	0.09	58	306.8	0.002	0.05	58	204.7	0.001
c5	0.03	72	290.8	0.001	0.02	72	204.7	0.000
<u>c6+</u>	<u>244.98</u>	112.7	309.1	<u>8.534</u>	<u>708.45</u>	112.7	204.7	<u>16.344</u>
total	681.68			9.600	729.96			16.361

total enthalpy

25.960

HYDROTREATING Heat Balance

rxeffluent vapor-liquid split at 300

rxeff	f	K300	f*K300	K+L/V	L/V=2		L/V=2.065	
					v	K+L/V	v	
h2s	1.54	4.70	7.24	6.70	1.08	6.77	1.07	
h2	414.94	29.60	12282.22	31.60	388.68	31.67	387.88	
c1	12.56	11.90	149.46	13.90	10.75	13.97	10.70	
c2	3.07	4.45	13.66	6.45	2.12	6.52	2.10	
c3	0.84	2.18	1.83	4.18	0.44	4.25	0.43	
c4	0.51	1.39	0.71	3.39	0.21	3.46	0.21	
c5	0.35	0.60	0.21	2.60	0.08	2.67	0.08	
<u>c6+</u>	<u>953.57</u>	<u>0.12</u>	<u>109.66</u>	<u>2.12</u>	<u>51.85</u>	<u>2.18</u>	<u>50.30</u>	
total	1387.38				455.21		452.77	
L/V					2.048		<u>2.064</u>	

rxeffluent vapor-liquid split at 400

rxeff	f	K400	f*K400	K+L/V	L/V=1.1		L/V=1.161	
					v	K+L/V	v	
h2s	1.54	7.00	10.78	8.10	1.33	8.16	1.32	
h2	414.94	22.80	9460.63	23.90	395.84	23.96	394.83	
c1	12.56	13.00	163.28	14.10	11.58	14.16	11.53	
c2	3.07	5.20	15.96	6.30	2.53	6.36	2.51	
c3	0.84	3.20	2.69	4.30	0.63	4.36	0.62	
c4	0.51	2.18	1.11	3.28	0.34	3.34	0.33	
c5	0.35	1.22	0.43	2.32	0.18	2.38	0.18	
<u>c6+</u>	<u>953.57</u>	<u>0.37</u>	<u>352.82</u>	<u>1.47</u>	<u>240.01</u>	<u>1.53</u>	<u>230.45</u>	
total	1387.38				652.45		641.78	
L/V					1.126		<u>1.162</u>	

HYDROTREATING
Heat Balance

rx effluent enthalpy at 300

rxeff	v	MW	H	MMBtu/Hr	I	MW	h	MMBtu/Hr
h2s	1.07	34	253.0	0.009	0.47	34	183.9	0.003
h2	387.88	2	815.6	0.633	27.06	2	183.9	0.010
c1	10.70	16	213.1	0.036	1.86	16	183.9	0.005
c2	2.10	30	250.9	0.016	0.97	30	183.9	0.005
c3	0.43	44	256.4	0.005	0.41	44	183.9	0.003
c4	0.21	58	238.2	0.003	0.30	58	183.9	0.003
c5	0.08	72	245.5	0.001	0.27	72	153.7	0.003
c6+	<u>50.30</u>	112.7	258.8	<u>1.467</u>	<u>903.27</u>	112.7	135.0	<u>13.743</u>
total	452.77			2.170				13.776

total enthalpy

15.946

rx effluent enthalpy at 400

rxeff	v	MW	H	MMBtu/Hr	I	MW	h	MMBtu/Hr
h2s	1.32	34	280.9	0.013	0.22	34	204.7	0.002
h2	394.83	2	1161.1	0.917	20.11	2	204.7	0.008
c1	11.53	16	279.9	0.052	1.03	16	204.7	0.003
c2	2.51	30	311.0	0.023	0.56	30	204.7	0.003
c3	0.62	44	317.8	0.009	0.22	44	204.7	0.002
c4	0.33	58	306.8	0.006	0.18	58	204.7	0.002
c5	0.18	72	290.8	0.004	0.17	72	204.7	0.003
c6+	<u>230.45</u>	112.7	309.1	<u>8.028</u>	<u>723.12</u>	112.7	204.7	<u>16.682</u>
total	641.78			9.051	745.60			16.705

total enthalpy

25.756

HYDROTREATING Heat Balance

rx effluent at dew point of 463

rxeff	f	MW	H	MMBtu/Hr
h2s	1.54	34	298.4	0.016
h2	414.94	2	1380.4	1.146
c1	12.56	16	324.6	0.065
c2	3.07	30	351.6	0.032
c3	0.84	44	358.5	0.013
c4	0.51	58	348.5	0.010
c5	0.35	72	337.0	0.008
<u>c6+</u>	<u>953.57</u>	<u>112.7</u>	<u>340.5</u>	<u>36.597</u>
total	1387.38			37.888

combined feed at dew point of 460

cfeed	f	MW	H	MMBtu/Hr
h2s	0.60	34	297.5	0.006
h2	441.55	2	1369.8	1.210
c1	12.56	16	322.5	0.065
c2	2.71	30	349.6	0.028
c3	0.60	44	356.6	0.009
c4	0.14	58	346.5	0.003
c5	0.05	72	334.8	0.001
<u>c6+</u>	<u>953.43</u>	<u>112.7</u>	<u>339.0</u>	<u>36.430</u>
total	1411.64			37.753

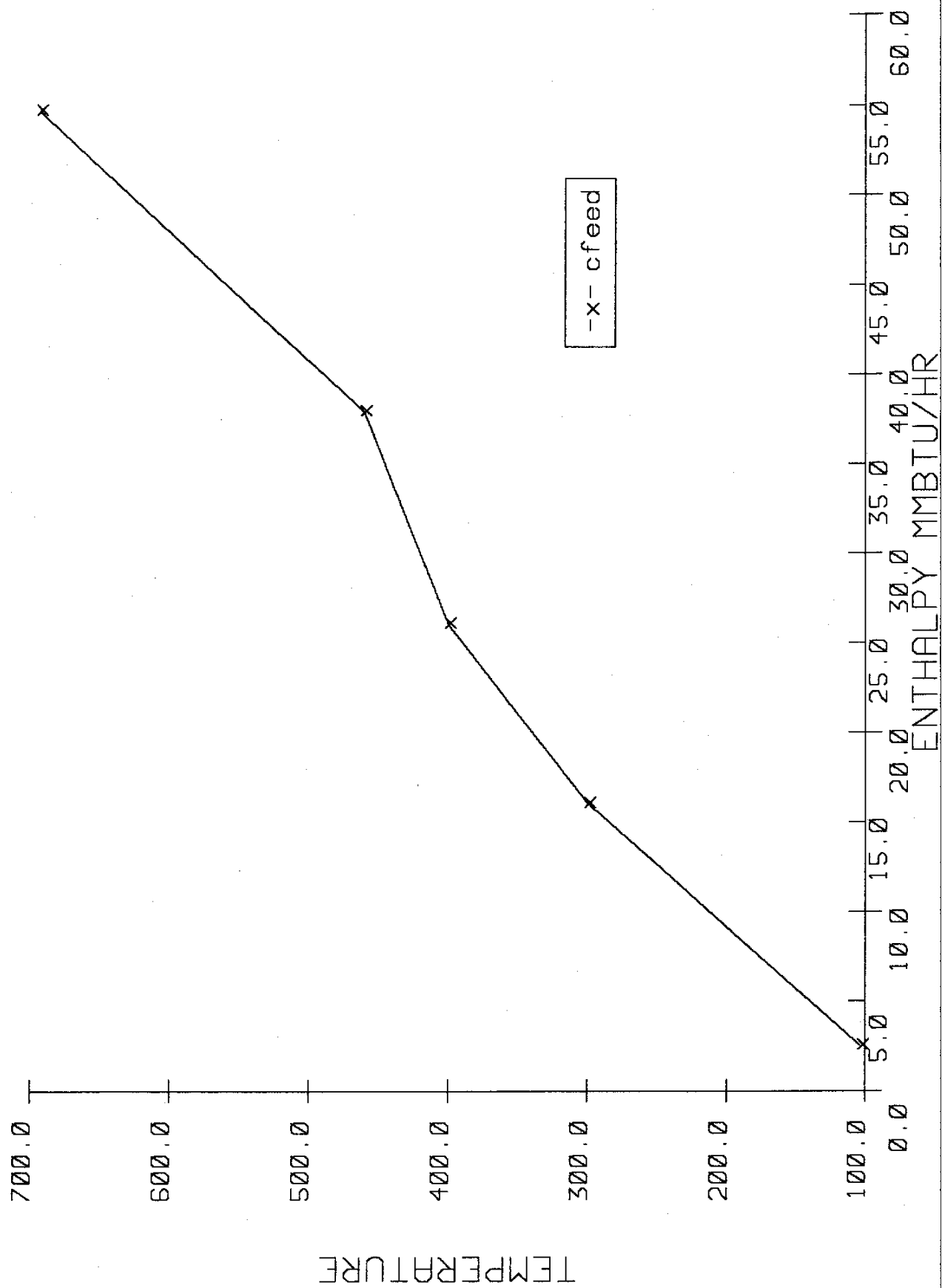
separator enthalpy at 120

sep	v	MW	H	MMBtu/Hr	I	h	MMBtu/Hr
h2s	0.60	34	216.4	0.004	0.94	36.6	0.001
h2	441.55	2	201.7	0.178	18.86	45.2	0.002
c1	12.56	16	111.1	0.022	3.43	45.2	0.002
c2	2.71	30	170.4	0.014	2.96	45.2	0.004
c3	0.60	44	182.1	0.005	1.96	45.2	0.004
c4	0.14	58	178.3	0.001	1.37	36.8	0.003
c5	0.05	72	179.0	0.001	1.08	33.5	0.003
<u>c6+</u>	<u>1.79</u>	<u>112.7</u>	<u>177.1</u>	<u>0.036</u>	<u>851.78</u>	<u>30.5</u>	<u>2.928</u>
total	460.00			0.261	882.38		2.947

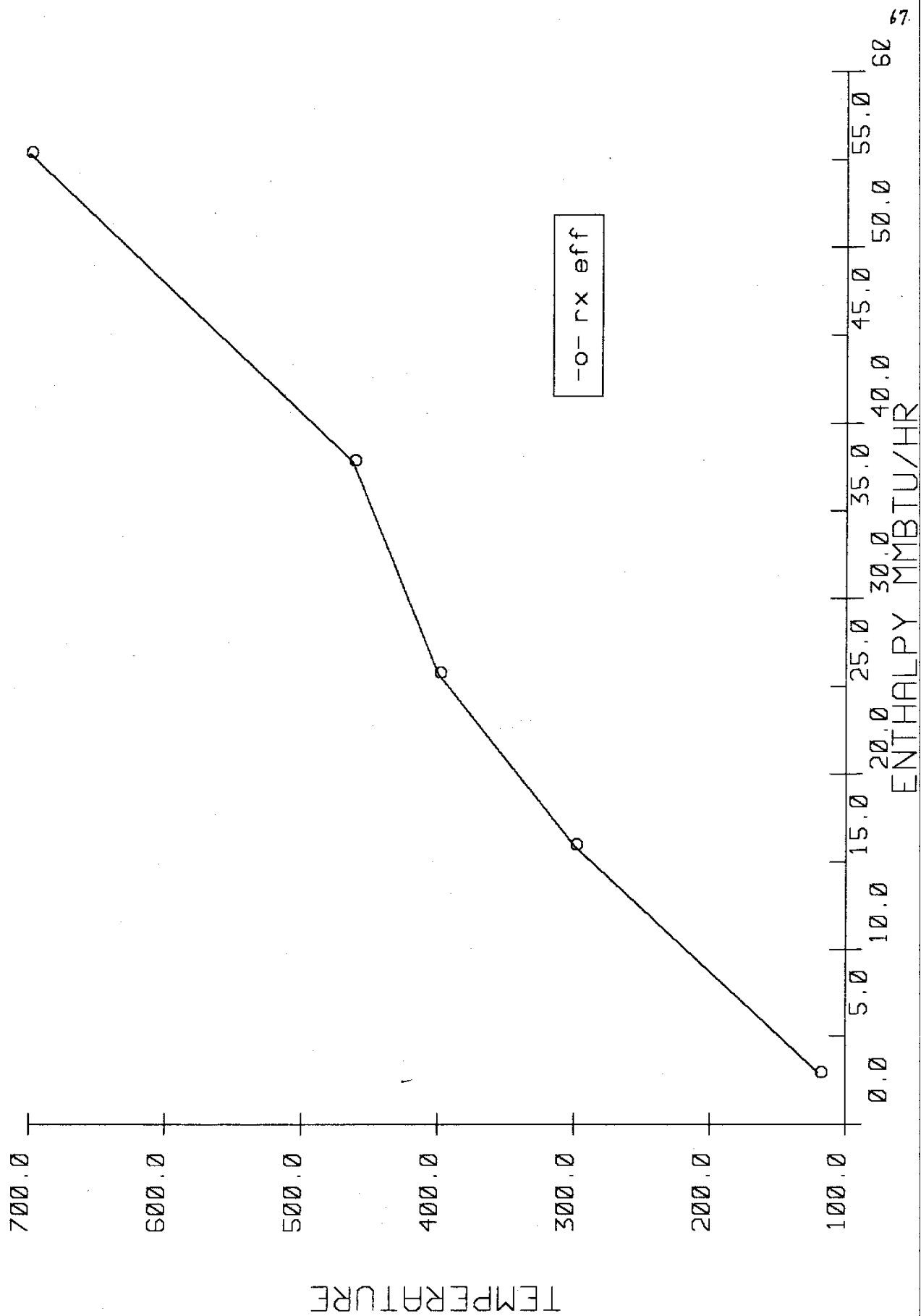
sep feed enthalpy

3.208

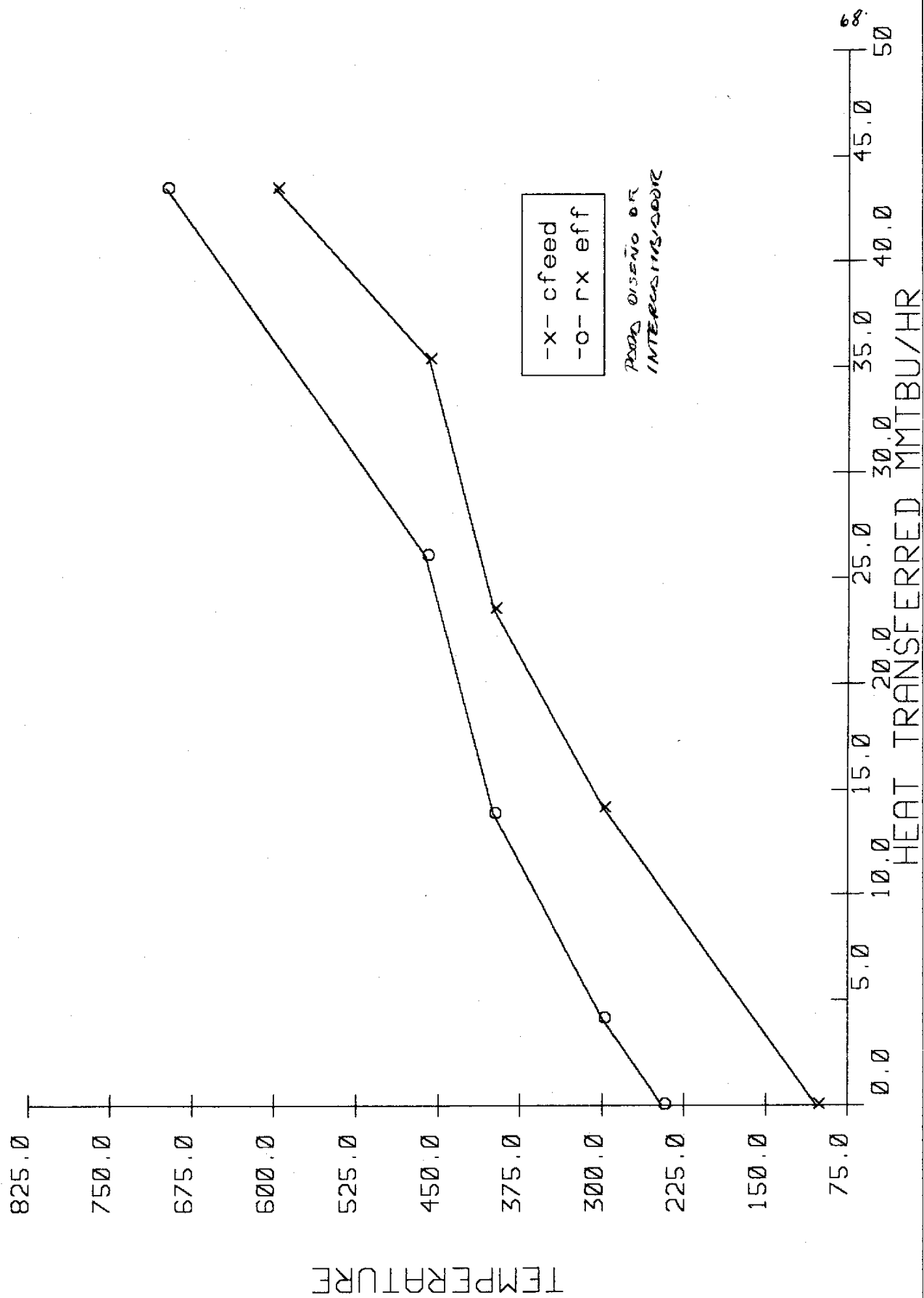
COMBINED FEED ENTHALPY

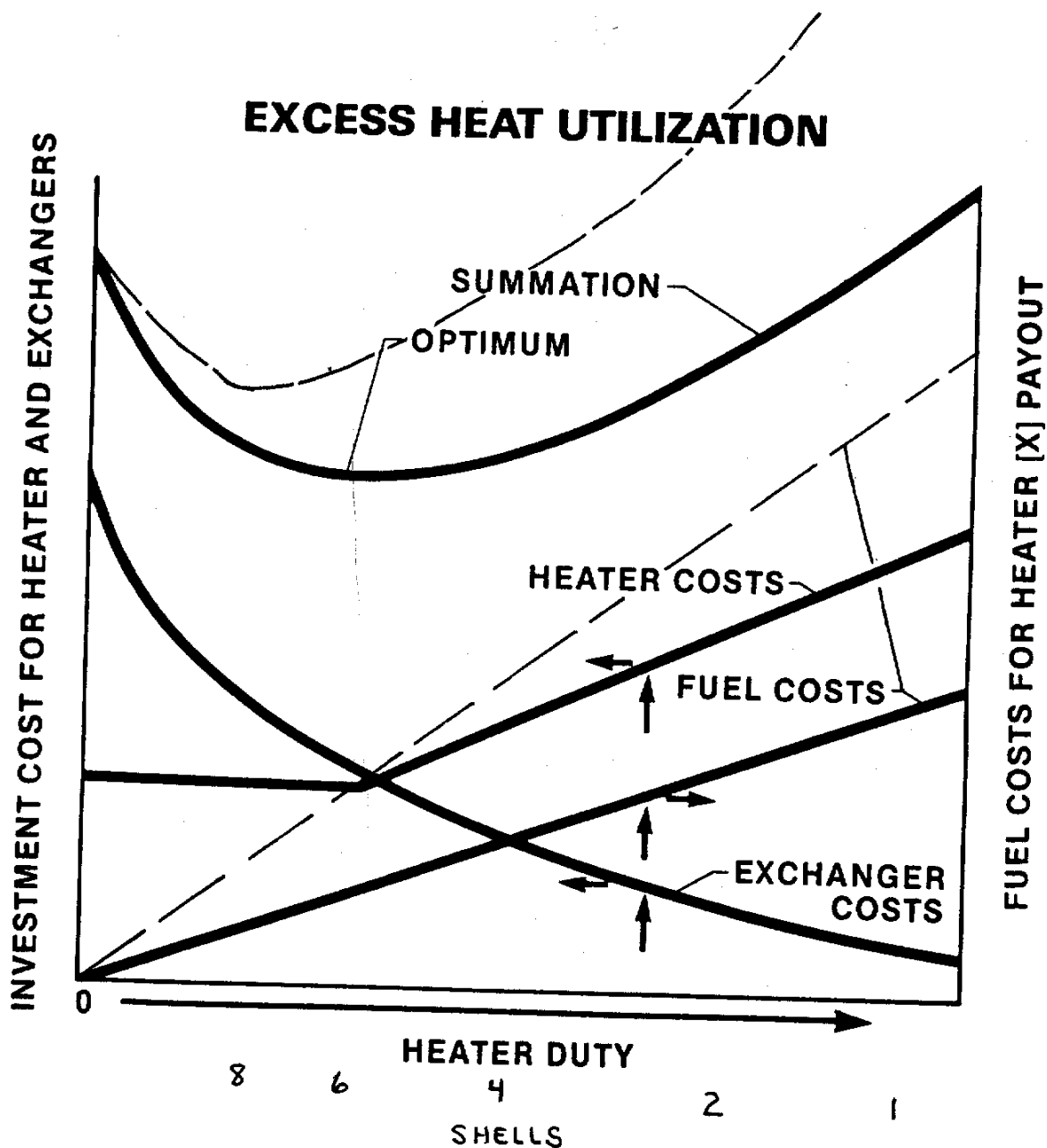


REACTOR EFFLUENT ENTHALPY



COMBINED FEED EXCHANGER





MINIMO TAMAÑO DEL HEATER.
6 SHELLS FOR EXCHANGER.

HYDROTREATING Heat Balance

	feed fr pump	rcy comp dischg	cfeed to exch	exch duty	cfeed to heater	heater duty	cfeed to reactor
mol/hr	951.64	460	1411.64		1411.64		1411.64
lb/hr	107250	1433	10863		10863		10863
enthalpy	2.133	0.418	2.551	43.467	46.018	8.632	54.65
press	500	500	500		460		410
temp	100	177	104				692

	reactor effluent	exch duty	effl fr exch	mkup h2	mtl to cond	cond duty	mtl to sep
mol/hr	1387.38		1387.38	55	1442.38		1442.38
lb/hr	108683		108683	415	109278		109278
enthalpy	55.35	43.467	11.883	0.064	11.947	8.739	3.208
press	400		360	360	360		350
temp	700		245	100	230		120

HYDROPROCESSING UNIT DESIGN

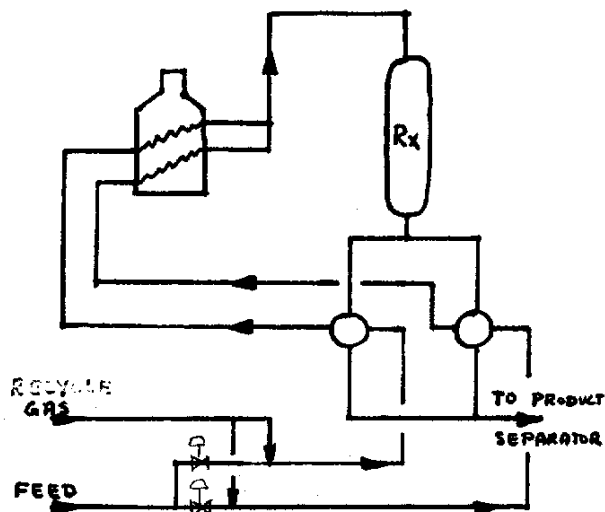
REACTOR CIRCUIT CONFIGURATION

HEATER/EXCHANGER SCHEMES

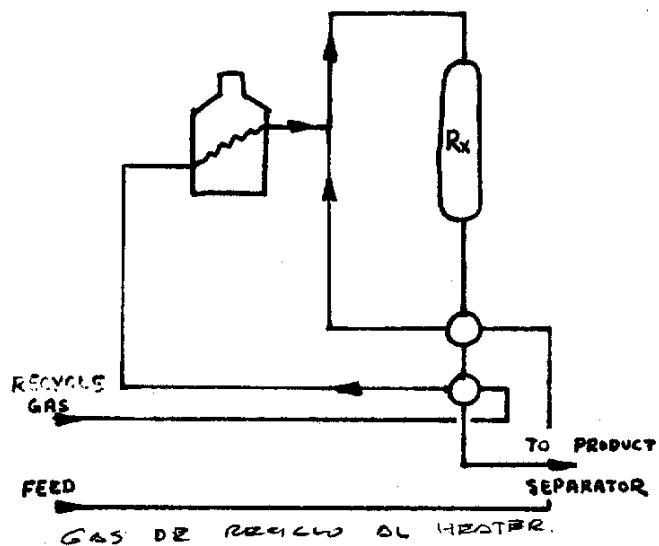
ALTERNATIVES

1. NORMAL FLOW SCHEME
2. THREE ALTERNATE FLOW SCHEMES
3. ADVANTAGES / DISADVANTAGES

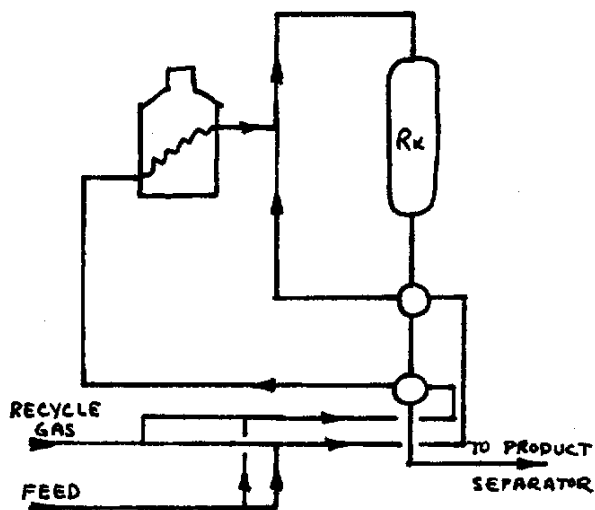
NORMAL FLOW SCHEME



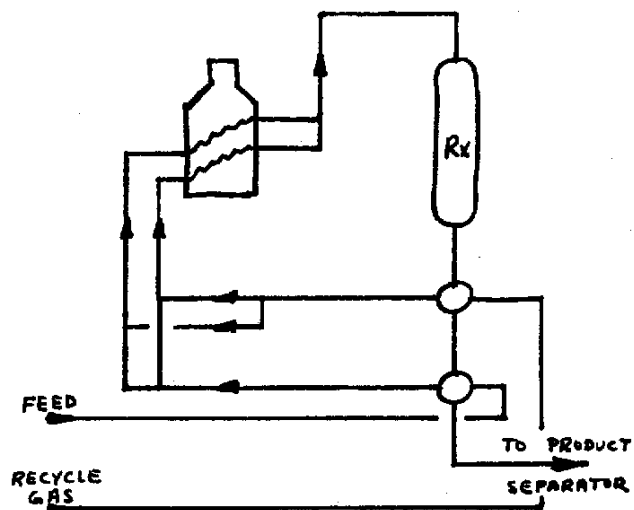
ALTERNATE SCHEME ONE



ALTERNATE SCHEME TWO



ALTERNATE SCHEME THREE



[illegible]

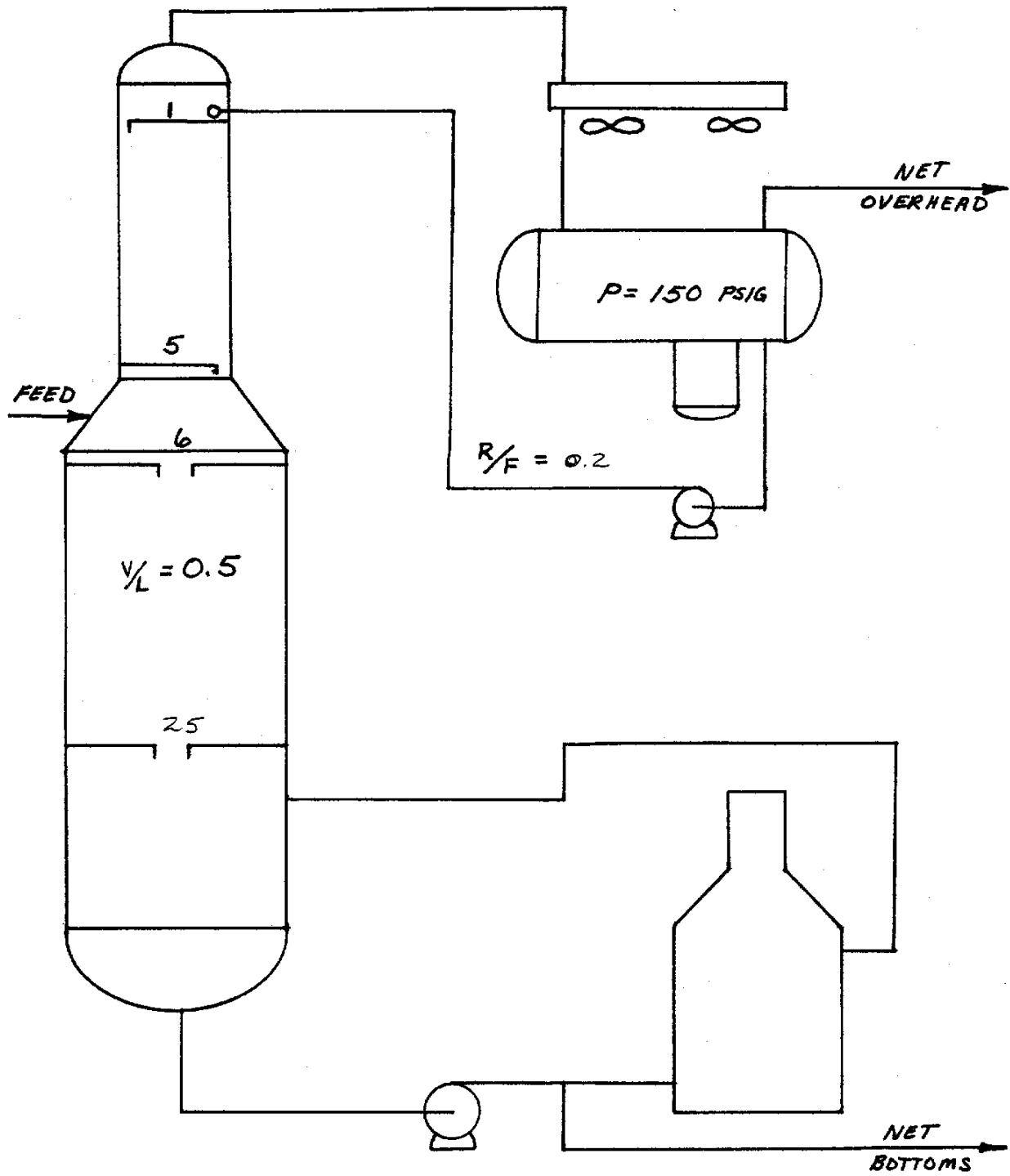
CONCLUSIONS

- 1) NORMAL FLOW SCHEME IS BEST FOR MOST UNITS
- 2) ALTERNATE SCHEME ONE IS USED FOR UNITS WHERE LIQUID FEED PROPERTIES WOULD RESULT IN EXCESSIVE CORROSION IN THE HEATER TUBES
- 3) ALTERNATE SCHEME TWO IS USED FOR UNITS WITH HIGH FEED RATES AND HIGH HEAT OF REACTION
- 4) ALTERNATE SCHEME THREE IS USED WHEN AVAILABLE PLOT SPACE PROHIBITS MULTIPLE EXCHANGER TRAINS

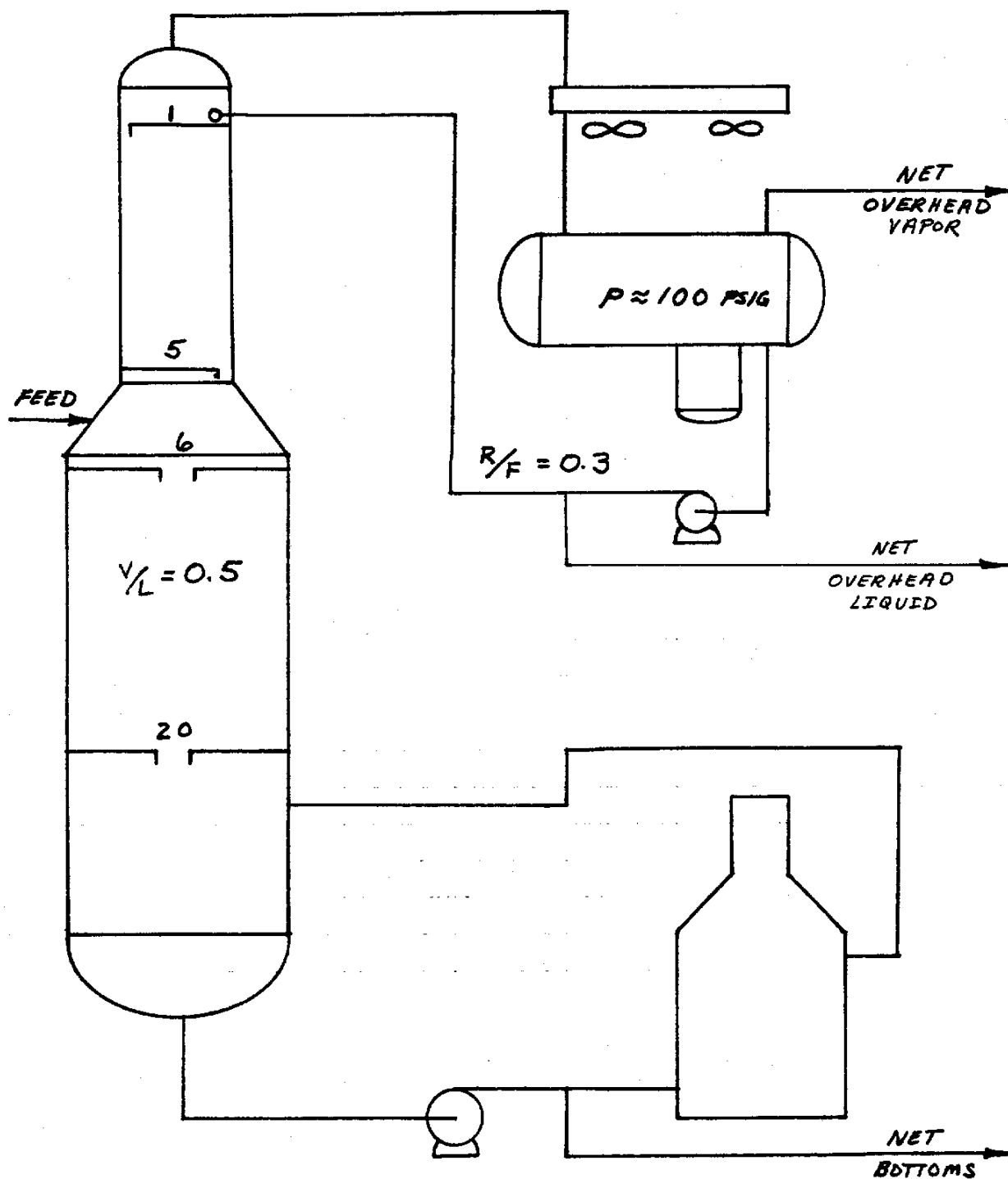
HYDROPROCESSING UNIT DESIGN
FRACTIONATION

- 1. NAPHTHA HYDROTREATERS*
- 2. KEROSENE HYDROTREATERS*
- 3. DIESEL HYDROTREATERS*
- 4. HEAVY OIL HYDROTREATERS*
- 5. HYDROCRACKERS*

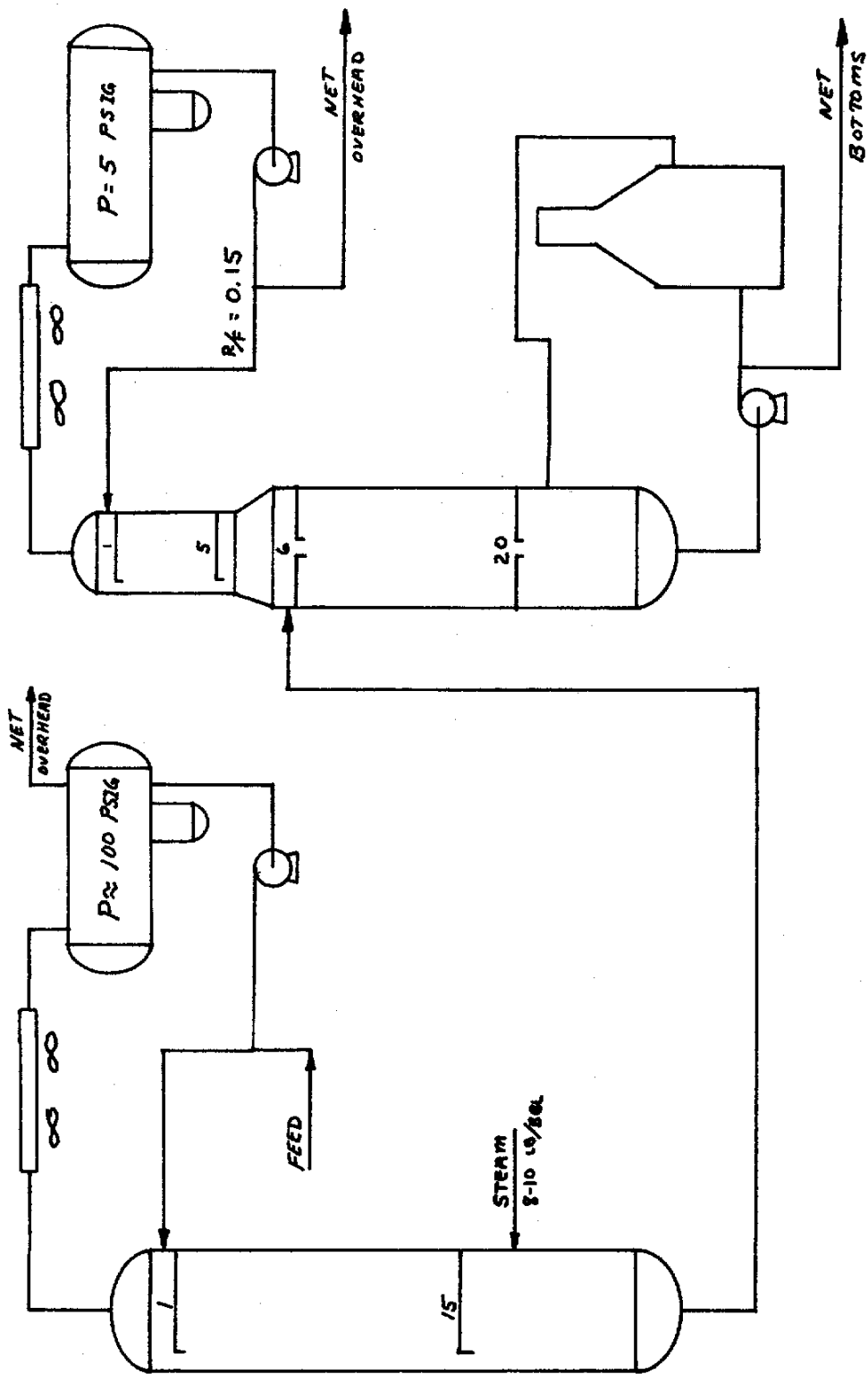
NAPHTHA HYDROTREATER STRIPPER



KEROSENE HYDROTREATER STRIPPER



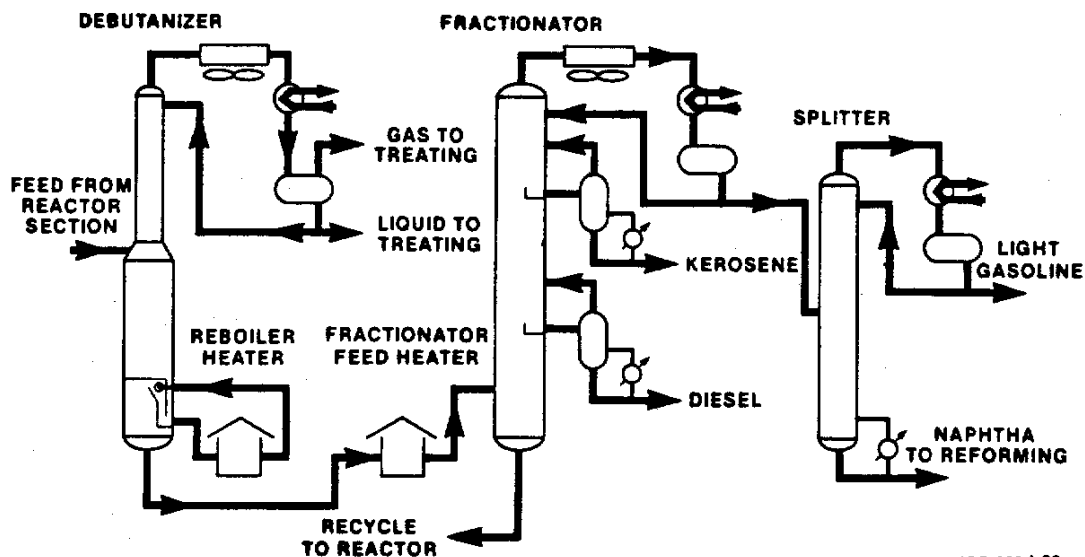
DISTILLATE HYDROTREATERS STRIPPERS



HYDROPROCESSING UNIT DESIGN

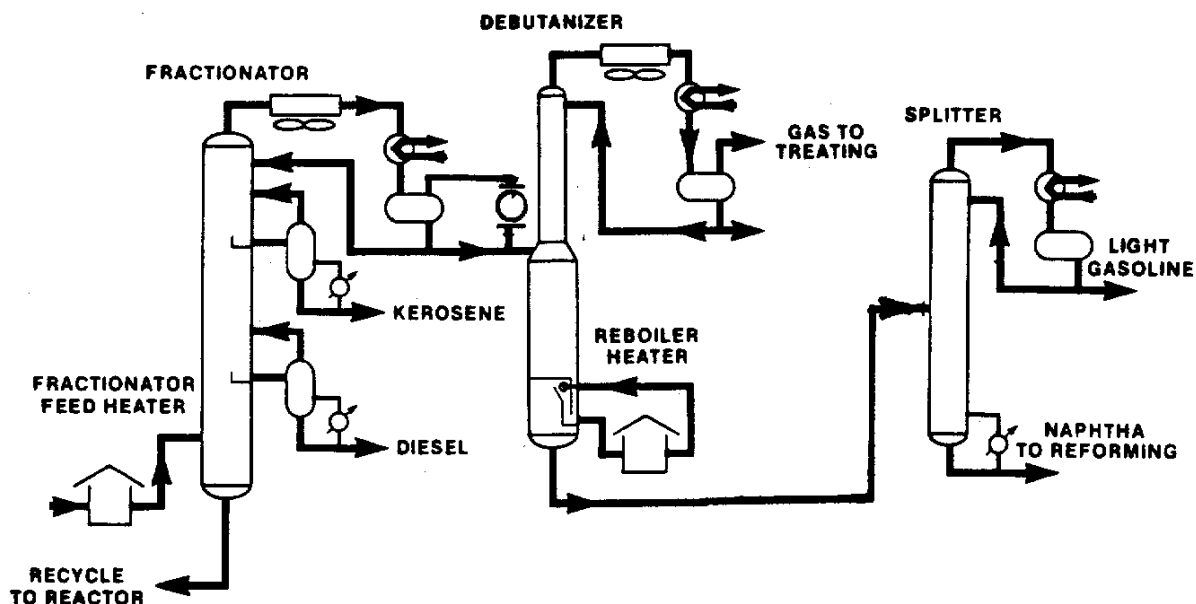
29

VI FRACTIONATION B. HYDROCRACKING



UOP 263J-26

- MINIMUM H_2S IN FRACTIONATOR
- NO SOUR GAS COMPRESSOR



- FRACTIONATOR METALLURGY UPGRADED
- DEBUTANIZER SMALLER

PROBLEM #1

What is the composition of the separator liquid and the recycle gas for the following hydrotreater? What is the make-up gas rate in moles/hr? The separator is to operate at 100°F and 350 psig.

Reactor Products (moles/hr)

H2	-40
H2S	10
C1	5
C2	8
C3	12
iC4	10
nC4	15
iC5	20
nC5	20
C6	Cut 1 100
	Cut 2 100
	Cut 3 100
	Cut 4 100
	Cut 5 100
Total	560

Make-up Gas (mole fraction)

H2	0.90
C1	0.05
C2	0.03
C3	0.02

PROBLEM #2

Using the separator liquid from Problem #1 and assuming perfect separation in the stripper, what is the composition of the reflux ~~rate~~ and the composition of the receiver vent gas?

Problem 3

Using the calculation method of problem one we find that the recycle gas purity for the following hydrotreater is 79.5 %. Since 80% has been specified by the yield estimate we must find a way to improve the recycle gas purity. If we achieve this with a hydrogen vent stream what is the resulting recycle gas composition, separator liquid composition, and make-up gas rate? Make-up gas rate with no vent is 61.9 moles/hr.

REACTOR PRODUCTS (moles/hr)		MAKE-UP GAS COMPOSITION (mole fraction)	
H ₂	-40	H ₂	0.80
H ₂ S	10	C ₁	0.10
C ₁	5	C ₂	0.06
C ₂	8	C ₃	0.04
C ₃	12		
i-C ₄	10		
n-C ₄	15		
i-C ₅	20		
n-C ₅	20		
C ₆ ⁺ cut 1	100		
cut 2	100		
cut 3	100		
cut 4	100		
cut 5	100		
TOTAL	560		

DATA FOR PROBLEMS

COMPONENT	EQUILIBRIUM $K's$	
	@ 350 PSIG / 100°F	@ 150 PSIG / 100°F
H ₂	51.94	137.80
C ₁	6.11	17.70
C ₂	1.61	4.19
H ₂ S	1.06	2.70
C ₃	0.69	1.28
i-C ₄	0.37	0.51
n-C ₄	0.28	0.37
i-C ₅	0.14	0.15
n-C ₅	0.11	0.12
C ₆ + out 1	0.04	0.04
out 2	0.01	0.01
out 3	0	0
out 4	0	0
out 5	0	0

Problem 1 solution

Assume
MUG =
50 moles/hr

Assume MUG = 55
Since H_2 is main
component and must
be approximately
doubled

Since K_i of other
components ~ constant
need K_i of H_2 = 0.836
try MUG = 55.45

	K_i	X_i	$K_i X_i$	X_i	$K_i X_i$	X_i	$K_i X_i$
H_2	51.94	$5/610$	0.426	$9.5/615$	0.802	$9.985/615.45$	0.836
H_2S	1.06	$10/610$	0.017	$10/615$	0.017	$10/615.45$	0.017
C_1	6.11	$7.5/610$	0.075	$27.5/615$	0.077	$27.723/615.45$	0.077
C_2	1.61	$9.5/610$	0.025	$9.45/615$	0.025	$9.664/615.45$	0.025
C_3	0.67	$13/610$	0.015	$13.1/615$	0.015	$13.109/615.45$	0.015
$i-C_4$	0.37	$10/610$	0.006	$10/615$	0.006	$10/615.45$	0.006
$n-C_4$	0.28	$15/610$	0.008	$15/615$	0.007	$15/615.45$	0.007
$i-C_5$	0.14	$20/610$	0.005	$20/615$	0.005	$20/615.45$	0.005
$n-C_5$	0.11	$20/610$	0.004	$20/615$	0.004	$20/615.45$	0.004
C_6 and 1	0.04	$100/610$	0.007	$100/615$	0.007	$100/615.45$	0.006
cut 2	0.01	$100/610$	0.002	$100/615$	0.002	$100/615.45$	0.002
cut 3	0	$100/610$	0	$100/615$	0	$100/615.45$	
cut 4	0	$100/610$	0	$100/615$	0	$100/615.45$	
cut 5	0	$100/610$	0	$100/615$	0	$100/615.45$	

0.583
≠ 1.0

0.966
≠ 1.0

0.999
close
enough

Recycle gas purity is 83.6% HYDROGEN

Makeup gas rate is 55.45 lb-moles/hr = 0.48 MMSCFD

Problem 2 Solution

	K_i	y_i	y_i/K_i	y_i	y_i/K_i	y_i	y_i/K_i	y_i
H_2	137.80	$\frac{2.95}{75.451}$	.0010	$\frac{2.95}{75}$	.0010	$\frac{2.95}{74.551}$	.0010	.133
H_2S	2.70	$\frac{10}{75.451}$	.0491	$\frac{10}{75}$	.0494	$\frac{10}{74.551}$	.0497	.134
C_1	17.70	$\frac{2.773}{75.451}$	.0058	$\frac{2.773}{75}$	.0059	$\frac{2.773}{74.551}$	.0059	.104
C_2	4.19	$\frac{9.664}{75.451}$	.0306	$\frac{9.664}{75}$	.0308	$\frac{9.664}{74.551}$	.0309	.130
C_3	1.28	$\frac{13.89}{75.451}$	.1357	$\frac{13.89}{75}$	.1366	$\frac{13.89}{74.551}$	.1374	.176
i-C ₄	0.51	$\frac{10}{75.451}$	.2599	$\frac{10}{75}$	.2615	$\frac{10}{74.551}$	.2630	.134
n-C ₄	0.37	$\frac{15}{75.451}$	.5373	$\frac{15.89}{75}$	.5243	$\frac{15.1}{74.551}$	.5112	.189
i-C ₅	0.15							
n-C ₅	0.12							
C_6^+ 1	0.04							
2	0.01							
3								
4								
5								

$\Sigma =$	$\frac{1.0194}{\approx 1.0}$	$\frac{1.0095}{\approx 1.0}$
		$\frac{0.9991}{\text{Close Enough}}$

Problem 3 solution

Assume make-up rate of 65 moles/hr

	SEPARATOR COMPOSITION	K_i	y_i trial 1	y_i trial 2	y_i trial 3
H_2	12.0	51.94	0.801	0.776	0.797
H_2S	10.0	1.06	0.017	0.017	0.017
C_1	11.5	6.11	0.110	0.109	0.110
C_2	11.9	1.61	0.031	0.031	0.031
C_3	14.6	0.69	0.016	0.016	0.016
iC_4	10.0	0.37	0.006	0.006	0.006
nC_4	15.0	0.28	0.007	0.007	0.007
iC_5	20.0	0.14	0.004	0.005	0.004
nC_5	20.0	0.11	0.004	0.004	0.004
ΣC_6+	100.0	0.04	0.006	0.006	0.006
2	100.0	0.01	0.002	0.002	0.002
3	100.0	0	0		
4	100.0	0	0		
5	100.0	0	0		
	625		1.004	0.979	1.000

Assume $V = 3$ moles and $L = 622$ moles (trial 1)

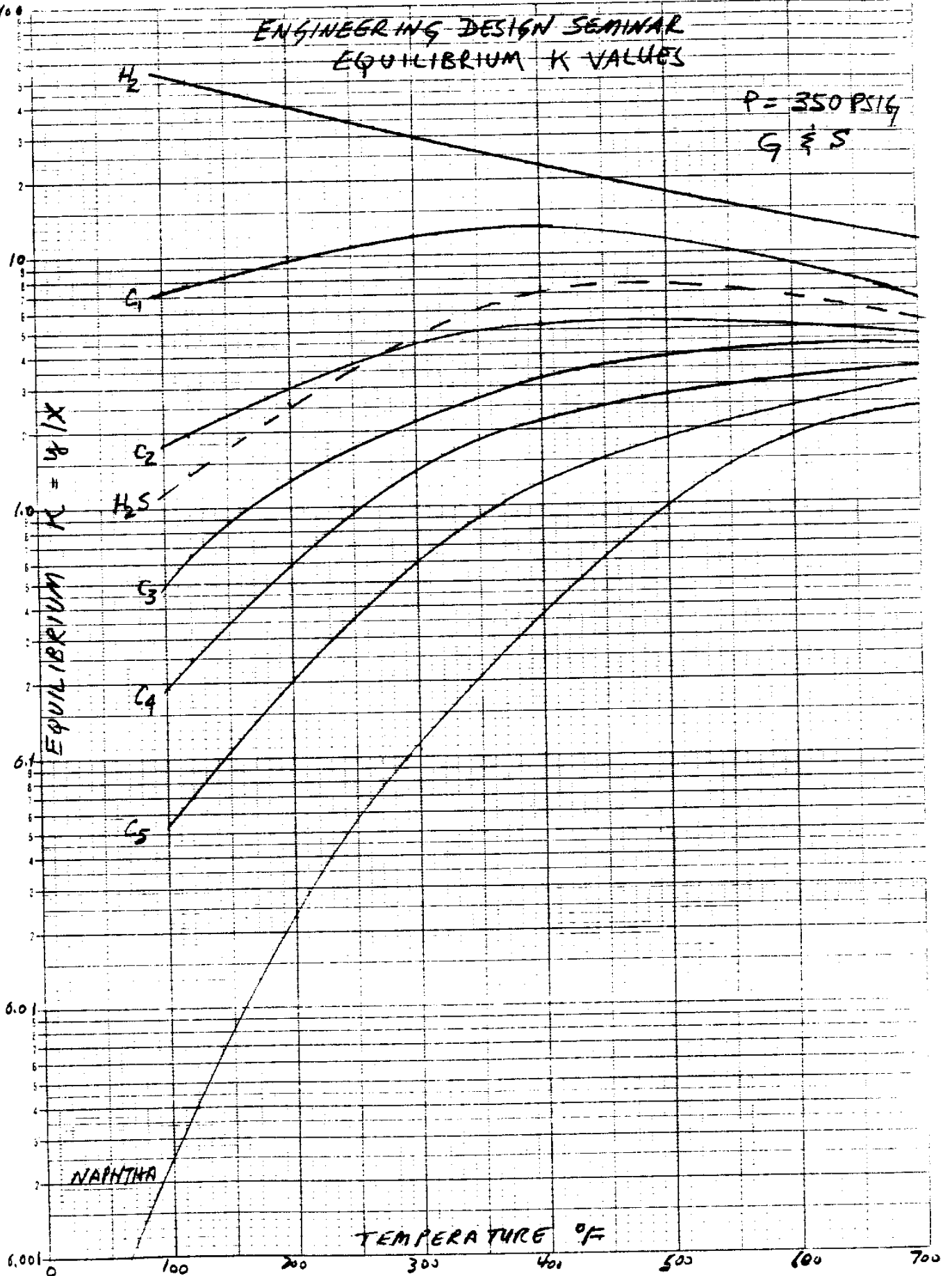
$$y_i = \frac{N_i}{\left(\frac{L}{K_i} + V\right)}$$

Assume $V = 3.5$ moles and $L = 628.5$ moles (trial 2)

Assume $V = 3.08$ moles and $L = 621.92$ (trial 3)

ENGINEERING DESIGN SEMINAR EQUILIBRIUM K VALUES

P = 350 PSIG
G = 5



problem -part 2 (cont'd)

additional design parameters

wash tower

assume that no makeup or withdrawal of wash oil is required.

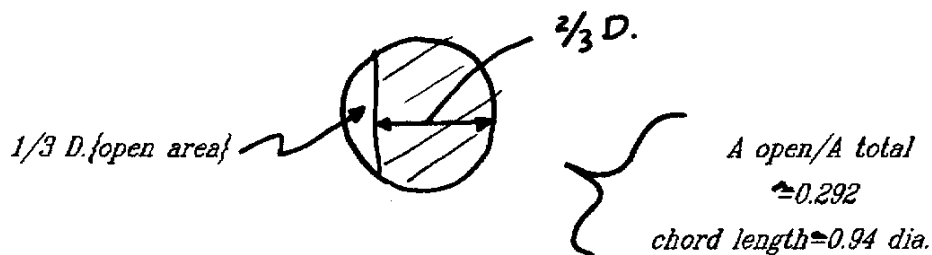
wash oil API is 20 UOP K is 11.7

max. allowable velocity in the wash tower = 2.0 ft/sec (0.61 M/sec)

-@ max. flue gas rate [assuming no internals in tower]

6 side-to-side pans in tower

(trayed area will cover 2/3 of the tower diameter, i.e.



K.O. Pot

max. allowance velocity in ft/sec

$$0.157 \sqrt{\frac{1}{\rho v}} = \text{velocity factor}$$

max flue gas rate $\div$ velocity factor = min allowance area of K.O. pot