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LBL-12115

CONF-810880--1

Lawrence Berkeley Laboratory
UNIVERSITY OF CALIFORNIA

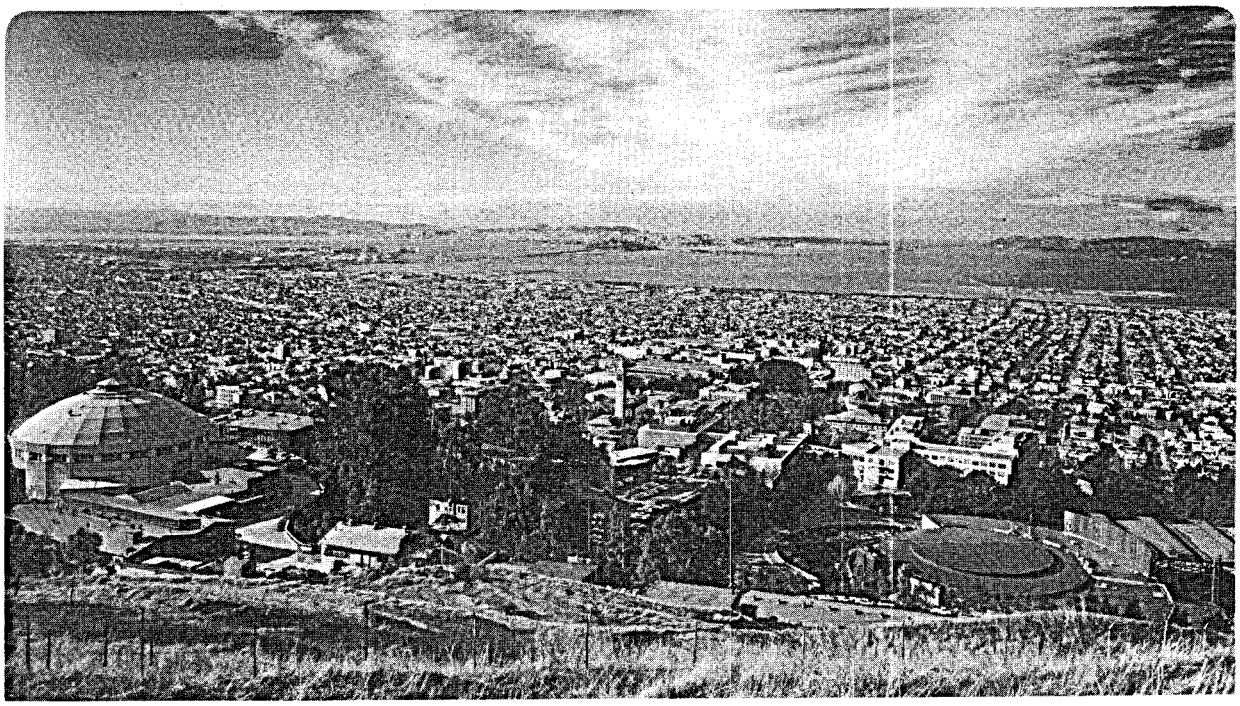
EARTH SCIENCES DIVISION

Presented at the 20th ASME/AIChE National Heat Transfer Conference, Knoxville, TN, August 2-5, 1981

DESIGN FEATURES AND EQUILIBRIUM FLASH MODELING OF DIRECT BINARY-FLUID HEAT EXCHANGERS FOR USE WITH GEOTHERMAL BRINES

Pascal M. Rapier

January 1981



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Prepared for the U.S. Department of Energy under Contract W-7405-ENG-48

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This work was supported by the Assistant Secretary for Conservation and Renewable Energy, Office of Renewable Technology, Division of Geothermal and Hydropower Technologies of the U. S. Department of Energy under contract no. W-7405-ENG-48.

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DESIGN FEATURES AND EQUILIBRIUM FLASH MODELING OF DIRECT-CONTACT BINARY FLUID HEAT EXCHANGERS FOR USE WITH GEOTHERMAL BRINES

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ABSTRACT

As part of the U.S. Department of Energy program to delineate and demonstration the economic advantages of the Direct-Contact Heat Exchange (DCHX) Process, Lawrence Berkeley Laboratory conducted baseline and preliminary tests of a 500 kW pilot plant during 1980. The DCHX evaporator, which combines an equilibrium-flash boiler with a countercurrent preheater, met and exceeded the performance of its predecessors, the volumetric and surface type boilers, although only its essential features had been installed. Additional tests to be conducted during 1981 are designed to improve overall tower performance and to create modular designs for a 20-foot diameter column, thereby replacing the plurality of DCHX units otherwise required for commercial-size (> 50 MWe) installations.

For the preheater-boiler combination, the recommended criterion of performance is the classical steam distillation efficiency, E_V, defined as the ratio of the partial pressure actually exerted by the hydrocarbon to its saturation pressure at the vapor exit temperature. As presently installed, the 500 kW DCHX unit typically generates a working-fluid vapor composed of 5 mole percent water and 95 percent isobutane at 415 psia and 254.2°F bubble point. E_V is 87.6 percent. However, with the additional improvements planned, 95 percent should be attainable.

INTRODUCTION

As part of the U.S. Department of Energy program to delineate and demonstrate the economic advantages of the Direct-Contact Heat Exchange (DCHX) Process, Lawrence Berkeley Laboratory (LBL) began testing a 500 kW DCHX pilot plant at its East Mesa site in January 1980. Emphasis was on accurately testing and modeling DCHX tower performance.

Preliminary operations of the DCHX tower during the 1980 fiscal year indicated that the available testing and modeling procedures were inadequate because these procedures assumed complete immiscibility of the hydrocarbon/water system, univariant 3-phase boiling, and separate preheater, boiler, and superheater units. These procedures did not create the correct preheater/boiler interface for the single DCHX unit actually installed; did not utilize the measured bulk thermal profile, which suggested strong backmixing down to the 12-foot level; and did not explain why no brine temperatures higher than the 2-phase boiling point were observed. These and other inadequacies indicated the need for an alternative to current univariant 3-phase boiling models.

PROCESS AND EQUIPMENT DESIGN

Equilibrium Flash Modeling of Heterogeneous Azeotropes

LBL had already recognized this requirement and had undertaken development of a partial miscibility systems analysis, a bivariant 2-phase boiling model, and an equilibrium-flash boiler/preheater combination unit [Rapier, 1978]. Two patent applications were made, one for the combination unit and the other for a multistage flash degasser. By removing CO_2 ahead of the DCHX boiler/preheater, the tendency for submerged boiling and off-gassing with the consequent backmixing is greatly lessened and the countercurrent operation of the preheater is improved.

LBL has derived phase diagrams for the isobutane/water system from data [Sage and Lacey, 1955] on the nearly immiscible system of n-butane/water by the principle of corresponding states. Using these mixture properties to replace pure component properties previously used with univariant 3-phase boiling models, LBL is employing a multi-component bivariant 2-phase boiling model which can closely duplicate the observed temperature profiles within the DCHX column.

Figure 1 shows that the process involves three steps: adiabatic mixing, isothermal separating, and isobaric preheating. While 3-phase models predict univariant azeotropic boiling, the observed boiling in the 500 kW plant is 2-phase with two degrees of freedom. The bubble point is about 12°F above azeotropic with about 5 mole percent water in the vapor phase at a pressure of 415 psia. The 12°F departure from azeotropic conditions is classically called the boiling point elevation (BPE) and at constant pressure depends uniquely on composition [McCabe and Smith, 1976]. Similarly, the composition of the vapor phase is 1.3 mole percent higher in water vapor than azeotropic, which is the cause of the BPE. The BPE of the mixture has either been overlooked in previous geothermal investigations or else mistaken for a superheat. Actually, the 2-phase boiling composition is a saturated dewpoint condition as shown by the phase diagram, Figure 2.

In active boiling, the isobutane fails to develop its full vapor pressure because of at least three factors contributing measureably to the total vapor pressure loss (VPL). These are:

a) Submergence, entrainment and demister losses.

b) Boiling point elevation (BPE).

c) Submerged nucleation and inert gas boil-off.

Minimization of all of these must be accomplished in the design of an efficient boiler/preheater. Submerged nucleation and offgassing of noncondensables disrupt the countercurrent flow and render the preheater less effective because of submerged volume boiling over portions of its height. Most important, however, is the boiling point elevation or vapor pressure depression that takes place because of imperfect equilibrium of the hydrocarbon liquid with excess water throughout the column. The two phase BPE is measureable as the rise in the bubble point temperature above the azeotrope as shown by the phase diagram for the working-fluid vapor composition. A brine BPE also occurs which is due to salt content and to demister losses. The total BPE sums the VPL losses and measures the overall degradation of the chemical potential for the process.

a) Adiabatic Feed Mixing Process. This process occurs in the entire froth and splash zone surrounding the hot brine injection nozzle. In the boiler, heat-transfer occurs only as a consequence of adiabatic flash-mixing and isothermal separation so that "pinch" temperature differences vanish and the concept of a thermal gradient is superfluous. The best boiler configuration is readily obtainable using conventional mixing nozzle and flash drum design, and staged equilibrium flash theory will yield all the necessary information for multi-component heat and mass transfer calculations. Volumetric heat-transfer coefficients for sizing the boiler mixing zone can be obtained from Figure 2 [Porter, 1965], as a function of the total mixing velocity at the brine nozzle.

b) Isothermal Separation of Working-Fluid Vapor and Residual Brine-Rich Liquid. The residual brine leaving the froth zone proceeds by gravity into the brine continuous region of the DCHX column. The working-fluid vapor rises out of the splash zone, and is demisted by a woven mesh demister. The working-fluid velocity through a horizontal demister should be near critical (flooding). For best performance, $V_c = 0.32\ ((\rho_L-\rho_G)/\rho_G)^{0.5}$ f/s.

c) Countercurrent DCHX Preheater Column. The preheater ideally functions as a countercurrent direct contact liquid-liquid heat exchanger from the top of the brine continuous zone downward to the extent that it is not disrupted by backmixing. The backmixing can occur from the brine nozzle above or by submerged boiling and offgassing of recirculated noncondensables from the IC_4 distributor below. The existence of countercurrent operation can be validated only by measuring actual temperature differences between the brine and hydrocarbon along the length of the column. Because of the boiler's equilibrium flash mixing operation, the brine and IC_4 streams merge; and the temperature difference approaches the brine BPE at the top liquid level of the column. Consequently, preheater "pinches" calculated from models which neglect boiling point elevations and the measured thermal profiles are inappropriate for determining the actual preheater performance.

DISCUSSION AND CALCULATIONS

BPE, LMTD and Volumetric Heat Transfer Coefficients

The brine boiling point elevation (BPE_b) is the summation of the BPE's due to salt and to transport pressure drops in the brine splash zone and in the demister. For 3000 ppm, the salt BPE is 0.1°F. At design vapor flowrates the transport BPE is about 0.3°F, so that the total brine BPE is 0.4°F for the 1980 series of runs.

LBL has calculated the minimum LMTD for the preheater by using the 0.4°F BPE_b as the TTD at the top of the column and the difference between the brine outlet and hydrocarbon inlet temperatures as the ITD at the bottom of the column. This is in accord with standard engineering practice in the seawater conversion industry for a countercurrent preheater.

The volumetric heat transfer coefficient for the preheater, U_v is defined herein as,

$$U_v = Q/vLMTD_{Min.}\ \text{btu/hft}^3\text{°F}$$

wherein v is the volume of the brine-continuous length of the DCHX column. With this definition, the U_v's are functions of accurately determinable process variables only.

Alternative Approach to Existing Models

LBL is also investigating using the measured bulk thermal profile within the preheater for deriving actual temperate profiles for the brine and isobutane, and for comparing ideal and actual preheater performance.

For example in Run 7, since about one-third of the 25-foot preheater length is active because of backmixing (compare in Table 1 Bulk ΔT/ft, Run #7 vs ideal), the bulk temperature rise per foot is only 27 percent that of an ideal preheater.

From this comparison of the volume ratio of the operating and ideal preheaters LBL defines the volumetric preheating efficiency E_p as the ratio of the mean respective bulk temperature rises per foot. In this way, the measured temperature profile of the column can be quantitatively taken into account.

Vaporization Efficiency

LBL's study of phase equilibria in DCHX operations reveals that the most important criterion for measuring performance and for improving design is the classical steam vaporization efficiency, defined as the partial pressure actually exerted by the hydrocarbon, divided by its saturation pressure at the vaporization temperature [Perry, 1950]. The baseline tests at design flow boiled off vapors at about 237.0°F, a pressure of 338 psia and a vaporization efficiency of 85 percent. During July, the boiling temperature was increased to 252.1°F at a pressure of 414.7 psia and a vaporization efficiency of 89 percent. It is expected that the two phase boiling temperature can be raised to 255°F and the pressure to 450 psia when 340°F brine becomes available and when a higher velocity brine nozzle is installed in the DCHX unit. The vaporization efficiency corresponding to these conditions is 94.88 percent, which is close to the 95 percent predicted by [Carey, 1930].

The highest efficiency results in a working-fluid having the highest pressure and lowest steam fraction for a given temperature. LBL has developed detailed phase diagrams for the above three operating pressures, so that the most accurate assessment of the operations can be made. Notice that the vaporization efficiency is a unique function of the VPL and BPE of the mixture.

The vaporization efficiency is calculated from the equations:

$$E_v = \left[p_{iC_4}/p^{sat}_{iC_4}\right]_{t_{vap}}$$

and

$$p_{iC_4} = \left[P_T - p^{sat}_{H_2O}\right]_{t_{vap}}$$

where t_{vap} = the temperature of the working-fluid vapor

P_T = the total pressure at the DCHX vapor exit

p_i = the partial pressure of the ith species

The power of this criterion can be visualized thus: suppose that the brine were flashed down to the equilibrium temperature 254.2F, and the flash steam were used in tubes to vaporize the IC_4. The same enthalpy as before would be transferred to the IC_4 vapor, but at a 12.4 percent greater pressure. The steam condensate, along with the flashed brine would be used for preheating the IC_4. Thus, one use of the criterion is to compare direct-contact and tubular exchange processes.

Data Evaluation

The data of 20 runs of the DCHX unit at East Mesa have been grouped into five averaged sets and analyzed. The results are summarized and presented in Table 1 along with those of three typical runs. Table 2 describes the various runs.

Average #3 of the most stable runs demonstrates a boiling point of 253°F at 415 psia for the working-fluid composition of 5 mole percent water and 95 mole percent isobutane. The vaporization efficiency, E_v, was 88.67 percent, the vapor pressure depression, VPL, was 49 psi, and the total boiling point elevation, (BPE) was 11.1°F.

Expressed in millions of btu's/hour, the heat-transfer rates were 10.12 for the preheater 7.13 for the boiler and 17.42 for the overall column. The respective volumetric heat-transfer coefficients expressed as UA/v, btu/ft^3°F hr were 3985 for the preheater, 38,000 for the boiler, including just the froth and brine mixing zone (top 2 feet of liquid), and 6880 overall (boiling plus preheating) for the 25 foot brine-continuous zone of the column.

Graphical Solution of Heat and Mass Balances

Run #7 from Table 1 is illustrated on the T-X diagram, Figure 2: at the nozzle, the hot flashing brine at 320.2°F and preheated IC_4 droplets are mixed adiabatically. As the brine flashes down to the equilibrium pressure of 414.7 psia it mixes with and vaporizes all the working-fluid, and leaves a residual brine-rich liquid. By conservation of mass, the preheating and mixing process operates solely along the vertical 0.8 mf water composition line, and the mixture boils continuously at 254.2°F in the 2-phase (L_2 + G) region of the phase diagram. At this temperature, azeotropic boiling occurs at 471 psia, so that there exists a 56 psi vapor pressure loss (VPL) in two phase boiling. This VPL is equivalent thermodynamically to a 12.2°F BPE, as shown by the T-X diagram.

To obtain a heat and mass balance by the [McCabe and Smith, 1976] method, an H-X diagram rather than a T-Q plot (which can be misleading) must be used. Referring to the H-X work sheet, Figure 4, and drawing a straight line from the brine inlet bubble point of 320.2°F so as to intersect the IC_4 rich liquid line at the bubble point of 254.2°F, and interpolating to the observed feed mass composition of 55 percent water by weight yields a brine temperature at the top of the DCHX column of 254.6°F. This demonstrates a boiling point elevation of about 0.4°F for the brine, attributable to salt and to transport losses, and allows calculation of the LMTD for the preheater. Other lines, drawn as shown in cooler sections of the preheater solve for the coldest bulk liquid temperature and for the median temperature. The heat and mass balance can also be calculated from the equations of Figures 3 and 4 as shown by Exhibit #1 for Run #7.

TABLE 1
Summary of DCHX Performance
Runs of 7/22/80 and 7/23/80 at East Mesa

CODE DCHX Runs 7/22&23/80	Av#1 (Linear T-L-Plot)	Av#2	Av#3 (Most Stable)	Av#4	Av#5	Run#1 (Highest Ev)	Run#5 (Lowest Ev)	Run#7 (Most Stable)	Calc. Ideal
Temp. °F Brine In	319.54	320.1	319.6	319.09	319.38	319.1	319.5	320.2	
Temp. °F Brine Exit	151.36	154.15	153.13	153.1	152.25	156	155.1	155	
Temp. °F HC in	94.96	97.08	94.5	95.95	95.15	98.5	97.3	93.9	
Temp. °F WF Out	252.52	253.45	253.08	252.92	253.23	251.2	256.4	254.2	
PSIA WF Out	415.5	415.95	415.05	415.77	414.6	417.1	415	415	
$p(H_2O)$ PSIA	31.18	31.68	31.48	31.39	31.56	30.46	33.33	32.10	
P_{sat}(HC) PSIA	430.29	435.09	432.58	431.92	433.19	424.53	446.3.	437.18	
E_v	.8932	.8832	.8867	.8899	.8842	.9108	.8552	.8758	.938
VPL(psi)	45.97	50.82	49.01	47.54	50.15	37.86	64.63	54.28	26.8
BPE °F (Boil Pt. Elev.)	10.52	11.45	11.08	10.92	11.23	9.2	14.4	12.2	6.5
Bulk ΔT °F/ft. DCHX	5.08	4.84	4.92	4.9	4.91	4.45	4.95	5.02	18.65
E_p	.272	.26	.264	.263	.263	.239	.265	.269	1
25'Prhtr. U_v BTU/°F ft^3 hr	4000	3940	3985	4040	4030	3950	3909	3875	
2'Blr. U_v BTU/°F ft^3 hr	39,100	36,000	38,000	37,400	37,200	45,700	29,400	34,400	
25'Tower U_v BTU/°F ft^3 hr	6920	6833	6880	7000	6970	6890	6856	6670	
Blr./Prhtr. LMTD °F	11.32	11.45	11.68	11.45	11.43	11.49	11.54	12.07	
Potential Flash °F	67.02	66.65	66.52	66.17	66.15	67.9	63.1	66.0	66.5
Brine/HC Mass Ratio	1.205	1.236	1.211	1.1864	1.2193	1.132	1.236	1.2199	
Brine FR Mlb/hr	103.05	104.36	105.78	103.81	105.67	99.81	104.33	105.75	
HC FR Mlb/hr	85.51	85.48	87.333	87.55	86.66	88.17	84.41	86.69	
Prhtr. ΔH MMbtu/hr	9.847	9.808	10.12	10.064	10.02	9.869	9.81	10.17	9.87
Boiler ΔH MMbtu/hr	7.18	7.20	7.346	7.13	7.29	7.345	7.39	7.33	7.35
DCHX ΔH_T MM btu/hr	17.03	17.01	17.47	17.42	17.31	17.21	17.2	17.49	17.22

TABLE 2

DESCRIPTION OF DCHX PERFORMANCE RUNS

Code	Av#1	Av#2	Av#3	Av#4	Av#5	Run#1	Run#5	Run#7
Date	7/22/80	7/22/80	7/23/80	7/23/80	7/23/80	7/23/80	7/22/80	7/23/80
Times	8:38 9:02 9:32 10:02 10:32	18:23 18:25 18:22 18:32	5:02 6:02 7:02 8:02	22:32 20:02 5:32 6:32 7:32 8:32	5:32 6:32 7:32 8:32	22:32	18:22	5:02
Operation	Stable	Unstable	Stable	Unstable	Stable	High E_v	Low E_v	Most Stable
Thermal Profile	Most Linear	Jagged	High Stepped	Low Stepped	High Stepped	Low Stepped	Unstable Jagged	High Stepped

NOTE: Most thermal profiles were double stepped, indicating submerged offgassing and/or backmixing. The profiles all indicated crossflow up to 6 feet above the IC_4 inlet distributor. In the unstable, or surging, operating mode, offgassing occurred at about 190°F. In the stable mode it occurred at about 235°F. Free CO_2 entering with the brine was estimated to be about 400 ppm.

CONCLUSIONS AND RECOMMENDATIONS

In the boiler, the only heat-transfer occurs as a consequence of adiabatic mixing and isothermal separation. The best boiler configuration is readily obtainable using conventional mixing nozzle and flash drum design, and equilibrium stage theory will yield all the necessary information for multicomponent heat and mass transfer calculations for the entire DCHX column independently of any volumetric heat transfer modeling assumptions.

a) Comparison of measured bulk temperatures for operating and ideal preheaters leads to a volumetric preheating efficiency defined herein as the ratio of the mean respective bulk temperature rises per foot. In this way, the measured temperature profile of the column can be quantitatively taken into account. A typical preheating efficiency E_p is 27 percent. LBL will continue work to correlate the measured bulk temperatures throughout the DCHX column with equilibrium stage theory.

b) The brine inlet nozzle should be redesigned to provide high horizontal velocities for froth-breaking and for atomizing the IC_4 droplets as they reach the adiabatic mixing zone; to impinge against the sidewalls of the vessel, thereby de-entraining and distributing the residual brine; and to prevent a downward velocity component that could result in backmixing, CO_2 release and submerged boiling.

c) CO_2 evolution by submerged release from recirculating IC_4 must be prevented, since it has been observed to disrupt and destablize the entire column operation. The best handling will be to remove it from the feed brine ahead of the DCHX.

d) Thermal circulation patterns (double eddies) must be prevented in open spray columns by insulation and by judicious use of vertical baffling and/or a few sieve plates.

EXHIBIT 1

Run #7 Heat and Mass Balance

1. Preheater: $(\dot{m}_w - \dot{m}_s)(h_w^{254.2} - h_w^{155}) = \dot{m}_I (h_I^{254.2} - h_I^{93.9})$

$$(1.2199 - .01634)(222.75 - 122.88) = 1(153.96 - 36.7)$$

$$(1.20356)(99.87) = 117.26$$

$$120.20 = 117.26$$

Balance = 97.55% Heat Loss = 2.45%

2. Boiler: $\dot{m}_w(h_w^{320.2} - h_w^{254.2}) = \dot{m}_v h_v^{254.2} - h_I^{254.2} - \dot{m}_s h_w^{254.2}$

$$1.2199\ (290.49 - 222.75) = 1.0634\ (235.65) - .01634\ (222.75) - 153.96$$

$$1.2199\ (64.74) = 239.5 - 3.6 - 153.96$$

Equilibrium Flash 82.63 = 81.90

H by Mass Transfer +2.6 +2.6
85.23 = 84.50

Balance = 99.144% Heat Loss = 0.86%

3. Overall Heat Transfer above 93.9°F base:

$$(\dot{m}_w - \dot{m}_s)(h_w^{320.2} - h_w^{155}) + \dot{m}_s(h_w^{320.2} - h_w^{93.9}) = \dot{m}_v h_v^{254.2} - \dot{m}_I h_I^{93.9} - \dot{m}_s h_w^{93.9}$$

$$1.20356\ (290.49 - 122.8) + .01634\ (290.49 - 62.5) = 239.5 - 36.7 - 1$$

$$1.20356\ (167.69) + .01634\ (227.97) = 201.8 \text{ (Actual Heat Loss = 0.21\%)}$$

$$201.82 + 3.725 = 201.8$$

From Brine 205.55 = 201.8 To IC_4

Balance = 98.18% Heat Loss = 1.82%

Corrected brine/IC_4 Ratio = 1.2199/1.0161 = 1.20 for Run #7

Subscript notation
- I = isobutane liquid
- s = steam mass fraction
- v = working fluid vapor
- w = brine or water, liquid

h_i = btu/lb of the ith component, ith state

$\dot{m}_i$ = mass rate of the ith component

Heat and mass transfer calculation for Run #7, East Mesa, July 23, 1980, 5:02 a.m., in btu's extracted per flowing pound of isobutane

e) One sieve plate should be placed 12 feet above the IC_4 distributor plate to help release any dissolved CO_2; another should be placed just below the brine inlet nozzle to aid in distributing the brine flow, to prevent backmixing, and to provide means for separate measurements of the brine and isobutane liquid temperatures. Even with this, the terminal temperature difference near the top of the column will be in the noise level because of the equilibrium stage operation of the boiler.

For the combined preheater boiler combination, the recommended criterion of performance is the classical steam distillation efficiency, E_V, defined as the partial pressure actually exerted by the hydrocarbon at the working-fluid vapor temperature, divided by the saturation pressure of the hydrocarbon at the same temperature. For run #7, this is 87.6 percent. However, with hotter brine, a well designed brine injection nozzle and good CO_2 removal, 95 percent should be attainable.

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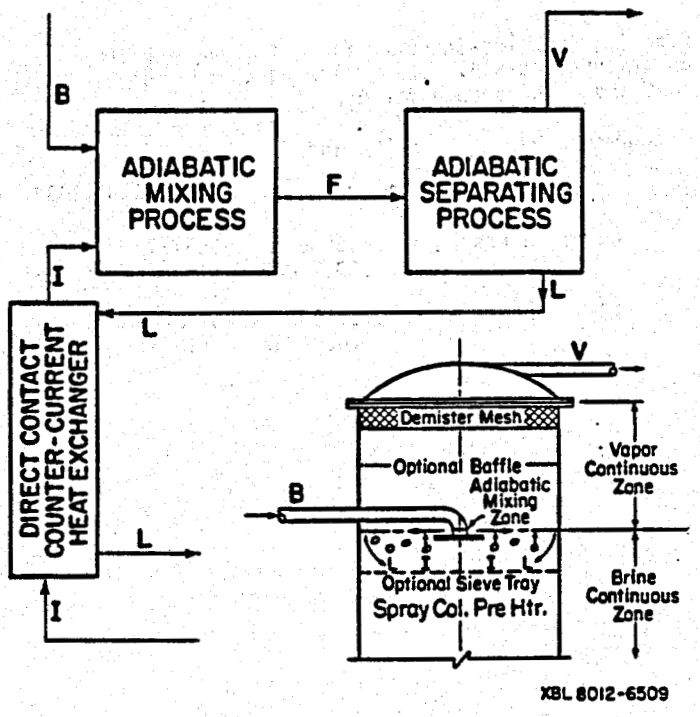

Fig. 1. Process block diagram for a direct-contact equilibrium flash boiler/preheater tower. The boiler comprises the vapor continuous zone, including the froth zone, while the preheater comprises the brine continuous zone.

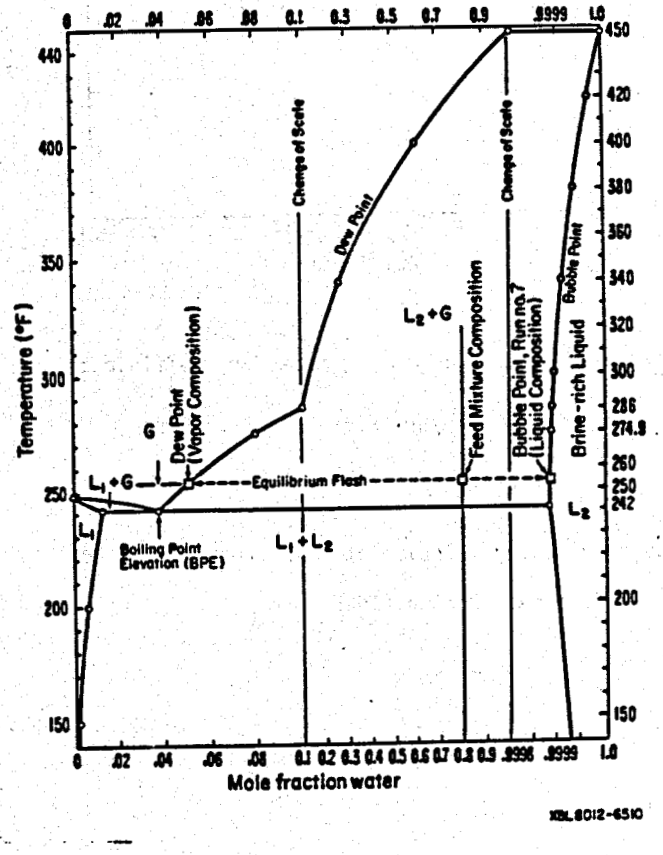

Fig. 2. Isobutane/water T-X phase diagram at P_T = 414.7 PSIA, showing equilibrium flash tieline, wherein the temperature along the dewline, $t(x)_{P_T}$ is related to the mole percent, x, by:

$$\mathrm{Ln}(\text{mol \% } H_2O) = 10.223965 - \frac{3136.973}{t(°F) + 110°}$$

and for an azeotrope vapor composition $(x_{AZ})_{P_T}$ the temperature and pressure, P_T, are related as

$$\mathrm{Ln}(\mathrm{PSIA}) = 12.819297 - \frac{4807.425}{T(°R)} + 5.3013E\text{-}39(T(°R))^{13}.$$

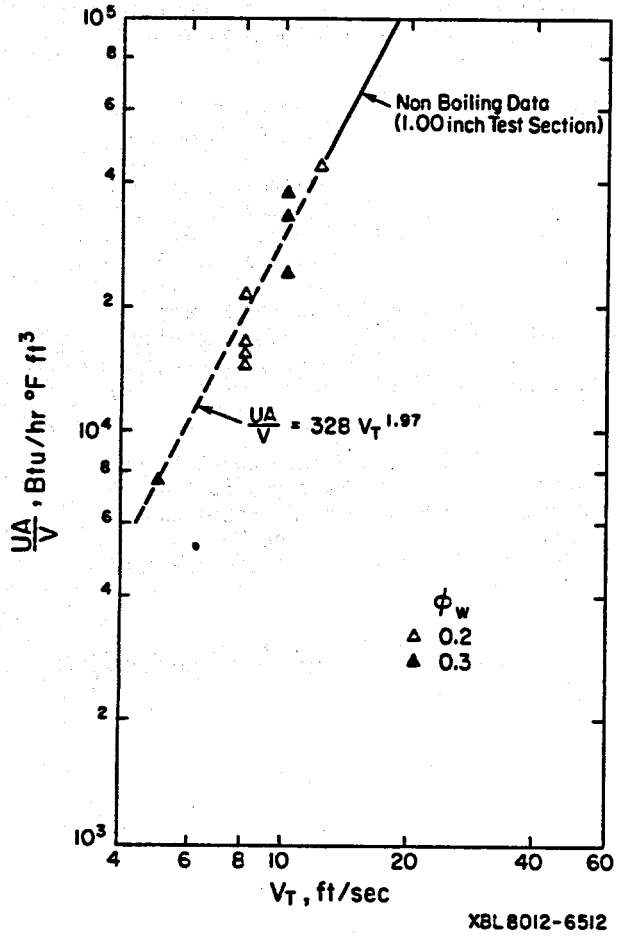

Fig. 3. Volumetric boiling heat transfer coefficients,

$U_V = \frac{UA}{v} = \frac{Q}{v\Delta T}$, btu/hr °F ft^3, ΔT = BPE

v = Volume of Mixer, ft^3

and V_T = mixing velocity at brine nozzle, ft/sec

Reference: Porter [1965]

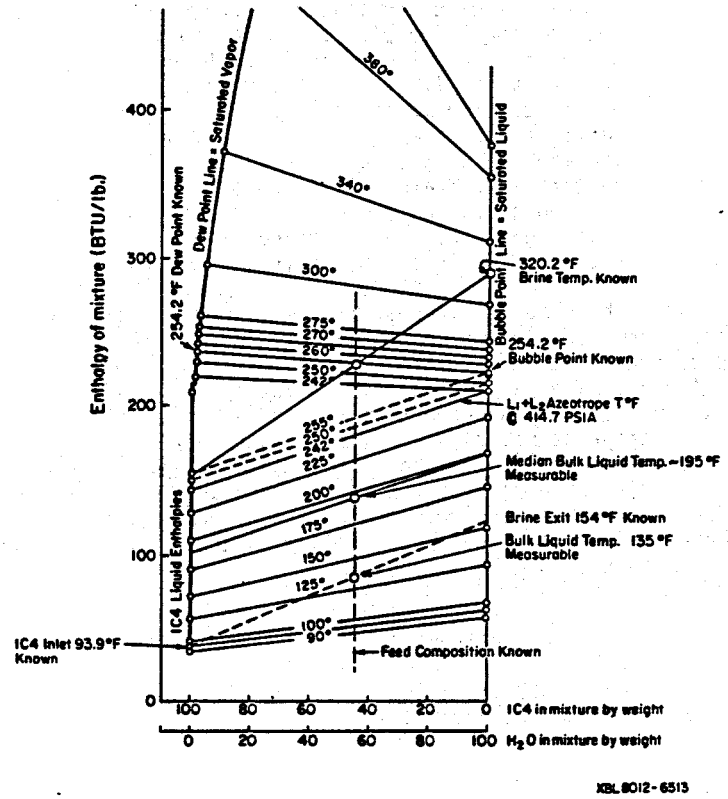

Fig. 4. Enthalpy-Concentration diagram, per lb of mixture, at 414.7 psia, showing the graphical heat and mass balance for Run #7,

where t = °F, and

dewpoints $h_V = 12.26641(t - 136.8)^{.620171434}$

bubblepoints $h_W = -28.50883 + .9583177t + .00011843t^2$ $(150 \leq t_W \leq 325)$

IC_4 droplets $h_I = -9.21296 + .399458t + .00095376t^2$ $(90 < t_{IC_4} \leq 255)$

This report was done with support from the Department of Energy. Any conclusions or opinions expressed in this report represent solely those of the author(s) and not necessarily those of The Regents of the University of California, the Lawrence Berkeley Laboratory or the Department of Energy.

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