Efficiency and Capacity Factors in Lube Extraction

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Premium-grade lubricating oils can be manufactured by phenol extraction of high-grade lube feed stocks. Aromatics of low viscosity index dissolve preferentially in the continuous phenol phase to form an extract, and paraffins and naphthenes of high viscosity index are relatively insoluble and form a discontinuous raffinate phase. Contacting is done in multistage plate-type towers using countercurrent flow between stages and cocurrent flow within stages. Phenol is recovered from both raffinate and extract for recirculation to the system. The process is very versatile and with relatively minor variations in process conditions is used to process stocks ranging from light napthenic base oils having a viscosity of 37 SSU at 210°F. to paraffinic premium-grade motor-oil bright stocks having a viscosity of 200 SSU at 210°F.

The results reported here were obtained during an experimental program aimed at increasing the contacting efficiency and capacity of the towers used in the extraction step. Factors considered are the effects of extraction temperature and interstage entrainment on extraction efficiency and of phase physical properties on capacity.

EFFICIENCY STUDIES

Definition of Extraction Efficiency

A simple, accurate measure of stage efficiency was needed to interpret the laboratory data and since some of the experimental work was done in a single-stage unit, the efficiency used had to be translated to multistage commercial towers.

Earlier work in these laboratories showed that refractive index (R.I.) was a good measure of lube-oil quality—viscosity index or paraffin plus naphthene content of the oil increasing with decreasing R.I. Thus the removal of aromatics is related to the decrease in R.I. of the raffinate oil as it passes through a stage. Efficiency measurements were therefore based on this inspection. Stage efficiency was defined as the ratio of the R.I. decrease actually obtained in the stage to that obtained by batch contacting to equilibrium at stage temperature, feed oil, and phenol in the same volume proportions as fed to the stage. This ratio was called "cocurrent" efficiency, defined as

$$\eta_c = \frac{I_F - I_R}{I_F - I_R^*} \tag{1}$$

Commercial-tower contacting ability is commonly expressed as over-all and/or Murphree efficiency. A relationship between these and the cocurrent efficiency was needed to relate results from the experimental single-stage unit to commercial extractors. The relation between Murphree and over-all efficiencies has been derived (*3*) and for lube extraction is given by

$$\eta_o = \frac{\log\left[1 + \eta_M\left(\frac{s}{m} - 1\right)\right]}{\log\left(\frac{s}{m}\right)} \tag{2}$$

The relationship between cocurrent and Murphree efficiencies is given by

$$\eta_M = \frac{\eta_c}{1 + \left(1 - \eta_c\right)\frac{s}{m}} \tag{3}$$

This expression in conjunction with Equation (2) relates the cocurrent and over-all efficiencies, a plot of the relationship being shown in Figure 1.

Equations (2) and (3) are based on straight operating and equilibrium lines. Study of a large number of lube extraction diagrams based on refractive index has shown that this requirement is fairly well approximated in many commercial extractions, at least over sections of the extractor.

Temperature Effect on Efficiency

Lube extraction plants generally process feed stocks of widely different viscosities in a single tower, and it was recognized qualitatively that extractor efficiency tended to be lower on high-viscosity stocks. This was studied to get a more quantitative effect of temperature and, hence, viscosity on extraction efficiency. Three lube-base stocks of different origin covering a wide viscosity range were used. Physical inspections on these oils are given in Table 1.

A single-stage extractor was used in the work, and a schematic diagram of it and the auxiliary

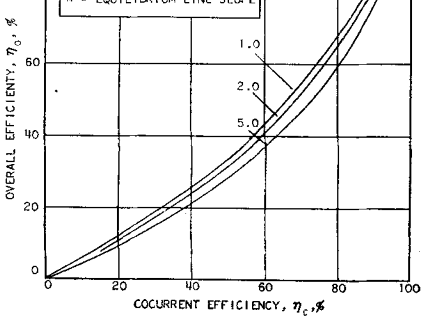

Fig. 1. Over-all and cocurrent efficiencies.

TABLE 1.—FEED STOCKS USED IN TEMPERATURE STUDIES

Feed stock	Gravity, °A.P.I.	Viscosity @ 100° F. SSU	Viscosity @ 100° F. Centistokes
Light Coastal	27.9	80	25.5
Canadian	25.7	160	34.3
Venezuelan	19.6	100 @ 210° F.	20.5 @ 210° F.

equipment is shown in Figure 2. The extractor was 4 in. thick with 20- by 30-in. glass front and rear faces with spotlights placed at the rear to permit observation of phase flows. Phenol was introduced to the stage through an internal downcomer; oil entered the stage just downstream of the downcomer outlet. Both phases then flowed through the mixing section of the stage and into the settling section. Phase mixing was caused entirely by the density difference between the phases. The vertical baffle in the mixing section and the oil-feed arrangement were the same in this series of experiments. Oil and phenol were fed to the unit with small gear pumps, flow rates being measured with calibrated rotameters. The oil was saturated with phenol prior to being fed to the unit to avoid the indeterminate effect which saturation within the stage would have on efficiency and to simulate partially operation of an intermediate tray in a commercial tower. Phenol was the continuous phase. The volume of oil holdup in the mixing section of the stage was measured by withdrawing a sample of total mixing-zone liquid through a sampler tube installed so that it might be swung across the mixing zone area at any stage height. These samples were allowed to settle at stage temperature and the fraction of oil phase was then obtained.

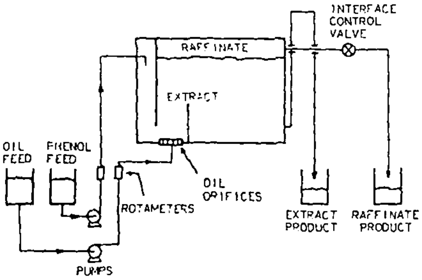

Fig. 2. Extraction equipment.

Table 2 lists the various temperatures and operating conditions used in the work together with phase physical properties and measured efficiencies. These data show that efficiency is increased by extracting at higher temperatures. With Venezuelan stock, for example, going from 139° to 200°F., at otherwise constant conditions, increased the efficiency nearly twofold. The effect of oil-phase viscosity on efficiency is illustrated by comparing the run on the Canadian oil at a volume flow ratio of 1.5 (line 2) with that on the Venezuelan oil at 139°F. (line 6). The change in raffinate viscosity from 7.7 to 31.5 centistokes caused a reduction in efficiency from 69 to 37%. The data illustrate also that the oil holdup in the mixing zone is an important variable. Comparison of the first two runs on the Canadian oil (lines 2 and 3) shows that increasing the oil holdup from 14 to 24% gave an increase in efficiency from 69 to 88%. The oil holdup was found to be dependent upon the oil rate, the volume flow ratio of phenol to oil, and the oil content of the extract phase.

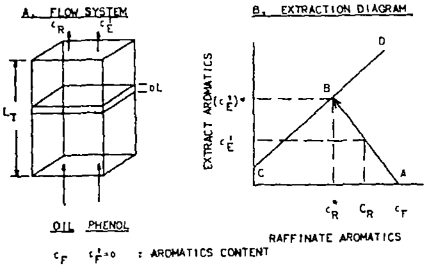

Fig. 3. Cocurrent extraction system.

To assist in understanding these results and to extrapolate them to conditions not directly measured, the extraction system was analyzed by assuming that back mixing of the oil and phenol did not occur in the mixing zone. Observation of the flow pattern in the cocurrent mixing zone indicated this to be approximately correct. It was assumed also that the extraction in the single stage could be represented by considering only the transfer of aromatics from the paraffins plus naphthenes to the phenol and that a straight equilibrium line existed over the limited concentration change in the stage. With these assumptions, the system can be represented by Figures 3 *a* and 3*b*.

With reference to Figure 3*a*, the aromatics transferred from raffinate oil to phenol in the length *dL* are given by

$$-R\,dc = K_R aA\,(c - c^*)\,dL \qquad (4)$$

Rearranging and integrating over the mixing-zone height L_T with constant K_R and R give

$$-\int_{c_F}^{c} \frac{dc}{c - c^*} = \frac{K_R aAL_T}{R} \qquad (5)$$

For a straight equilibrium line, *CD* in Figure 3*b*,

$$c = mc^* + b_1 \qquad (6)$$

By an aromatics material balance from the stage inlet to height L,

$$c' = (R/E)\,(c_F - c) = s\,(c_F - c) \qquad (7)$$

where s is the slope of the operating line *AB* in Figure 3*b*.

For oil drops of average diameter D and a fractional oil holdup in the mixing section of f the drop surface area per unit volume of mixing zone is

$$a = \left(\frac{6f}{D}\right) \qquad (8)$$

It was assumed further that the major mass transfer resistance was in an oil film at the surface of the drop, and so the over-all mass transfer coefficient K_R was equal to the oil film coefficient k_R. This latter coefficient, in turn, was considered to be related to the oil-phase physical properties by a power function of the form:

$$k_r = b_2\,(\nu_o)^{b_3} \qquad (9)$$

This simplification is limited to extraction systems having the same general physical properties, for example interfacial tension. Substitution of Equations (6) through (9) into Equation (5) gives

$$-\int_{c_F}^{c_R} \frac{cd}{c\left(1+\frac{s}{m}\right)-\left[\left(\frac{s}{m}\right)c_F - b_1/m\right]} = \frac{6b_2 fAL_T\,(\nu_o)^{b_3}}{DR} \qquad (10)$$

The integral in Equation (10) can be evaluated to give a function of the cocurrent efficiency η_c. Direct integration of the left-hand side of Equation (10) gives

$$\int_{c_F}^{c_R} f(c)\,dc = \left(\frac{1}{1+\frac{s}{m}}\right) \ln\left[\frac{c_F + (b_1/m)}{c_R + (b_1/m) - \left(\frac{s}{m}\right)(c_F - c_R)}\right] \qquad (11)$$

From Figure 3*b*

Table 2.—Viscosity Effect on Efficiency

Feed stock	Extraction conditions Oil rate, bbl./hr./sq. ft.*	Temp., °F.	Mix.-zone oil holdup, %	Vol. flow ratio, phenol to oil	Phase viscosity, centistokes† Raffinate	Extract	Cocurrent efficiency, %
Light Coastal	15.9	78	14	1.0	12.8	6.8	53
Canadian	15.9	141	14	1.5	7.7	3.1	69
	15.9	141	24	0.5	7.7	3.1	88
	5.0	145	7	1.0	7.2	2.9	79
	16.5	150	28	1.0	6.7	2.7	91
Venezuelan	11.9	139	12	1.0	31.5	3.0	37
	11.9	155	12	1.0	18.2	2.4	44
	11.9	200	14	1.0	9.1	1.3	70

* Based on mixing zone area.
† At extraction temperature.

$$\overset{*}{c}_E = m\overset{*}{c}_R + b_1 = s\,(c_F - \overset{*}{c}_R)$$

or

$$\frac{s}{m} = \frac{\overset{*}{c}_R + \frac{b_1}{m}}{c_F - \overset{*}{c}_R}$$

and

$$\eta_c = \frac{c_F - c_R}{c_F - \overset{*}{c}_R}$$

From these definitions of s/m and η_c there results

$$\left(1 + \frac{s}{m}\right) = \frac{c_F + \frac{b_1}{m}}{c_F - \overset{*}{c}_R} \qquad (12)$$

$$\left(1 + \frac{s}{m} - \eta_c\right) = \frac{c_R + \frac{b_1}{m}}{c_F - \overset{*}{c}_R}$$

Hence if the ln term in Equation (11) is divided through by ($c_F - c_R^*$), and Equation (12) and the definition for η_c are substituted there results:

$$\int_{c_F}^{c_R} f(c)\,dc = \left(\frac{1}{1+\frac{s}{m}}\right)\ln\frac{1}{1-\eta_c} \qquad (13)$$

Substituting (13) into (10) and rearranging to solve for η_c give

$$\eta_c = 1 - e^{\frac{-b_4\left(1+\frac{s}{m}\right) f A L_T (v_o)^{b_3}}{R}};$$

where

$$b_4 = \frac{6b_2}{D} \qquad (14)$$

Equation (14) is the result used in interpreting the experimental data in Table 2. Superficial oil contact time is given by fAL_T/R. Hence directionally a decrease in oil rate, an increase in mixing-zone area or height, or an increase in the fractional holdup of oil will increase extraction efficiency. As pointed out above, additional studies have shown that for lube-phenol systems, the fractional oil holdup f depends on the oil and phenol flow rate and the oil content of the phenol phase. Since f and c_3 depend on the physical properties of the extraction system and the dispersion method used, quantitative results from Equation (14) will depend on the specific physical and mechanical system under study; however, the general approach used to interpret the data is believed to be generally applicable and could be applied to other extraction systems.

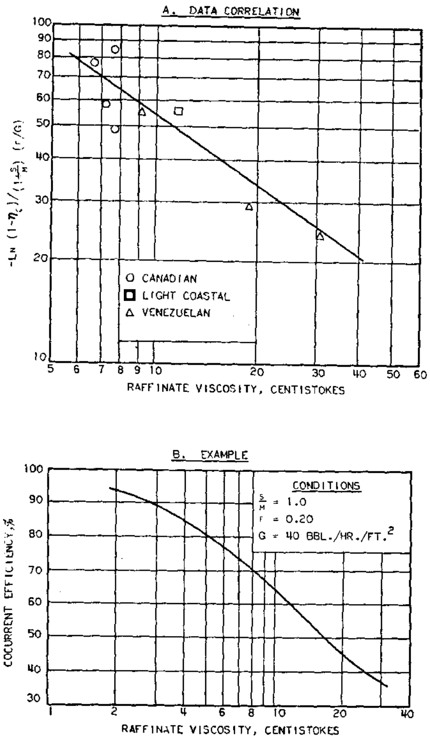

Fig. 4. Efficiency increased by reducing oil-phase viscosity.

Rearrangement of (14) gives

$$-\frac{\ln(1-\eta_c)}{\left(1+\frac{s}{m}\right)(f/G)} =$$

$$(b_4 L_T)\,(v_o)^{b_3} = b_5\,(v_o)^{b_3} \qquad (14a)$$

where G = the oil flow rate per unit of mixing-area cross section. By Equation (14a), a plot of the left-hand side vs. μ_o on log-log paper should yield a straight line. Figure 4a shows the data of Table 2 plotted in this manner. The agreement of the data with theory is considered satisfactory in view of the assumptions made in the analysis. That oil-phase viscosity may control in lube extractions is indicated by this analysis as the efficiency data of Table 2 are not correlated by the efficiency function if plotted vs. the phenol-phase viscosity.

The large effect of temperature and hence viscosity on efficiency is shown more clearly in Figure 4b. This plot was prepared for an s/m value of unity, a mixing-zone oil holdup of 20%, and an oil-flow rate of 40 bbl./hr./sq.ft. In the range of oil-phase viscosities greater than 4 centistokes, a 50% reduction in viscosity for these conditions will increase cocurrent efficiency by about 16%, for example, from 50 to 66% efficiency.

Interfacial tension between phases also affects stage efficiency by affecting drop size. With lube-phenol systems the interfacial tension is relatively constant at about 2.5 to 4.0 dynes/cm. and so oil drop diameters are relatively constant at about 1/16 to ⅛ in. To extend the range of interfacial tension, phenol was extracted from kerosene with water. Table 3 shows the extraction efficiency obtained in comparison with that for a typical lube-phenol system under comparable extraction conditions and with the same extractor. The relatively low lube efficiency was due to the use of a stage arrangement somewhat different from that used in the temperature-efficiency work.

The large drop size and low mixing-zone oil holdup caused by the high interfacial tension of the kerosene-water system reduced the available surface area and holding time more than enough to counterbalance the greater turbulence in the large kerosene drops. Efficiency was therefore reduced.

TABLE 3.—EXTRACTION EFFICIENCY REDUCED BY HIGH INTERFACIAL TENSION

System	Lt. Coastal oil-phenol	Kerosene-water
Oil viscosity, SSU @ 80° F.	69	33
Interfacial tension, dynes/cm.	3.1	17.0
Mixing zone oil holdup, %	17	6
Oil drop size (approx.), in.	1/16-1/8	1/4-1/2
Cocurrent efficiency, %	44	25

EFFECT OF ENTRAINMENT ON EFFICIENCY

In any extraction, entrainment of internal phases occurs if the settling capacity of the stage is exceeded. This becomes critical in certain lube extractions where the phenol and lube stock form slow separating emulsions. The entrainment will affect the contacting efficiency of the stages.

A brief literature review showed no estimate for the effect of dual-phase entrainment on efficiency in liquid extraction. Accordingly, an expression was derived to estimate the relative effect of raffinate phase entrainment vs. extract phase entrainment and to set some limit on the amount of entrainment to be tolerated in an extraction without seriously affecting over-all tower efficiency.

The following major assumptions were made in deriving the result:

1. The equilibrium and operating lines are linear.
2. Equal entrainment occurs in all internal stages.
3. The stages operate at 100% efficiency.
4. Oil solubility is constant in each phase.

The extraction system assumed in the derivation is shown in Figure 5. The normal raffinate flow from stage n is accompanied by entrained extract and recycle raffinate carried into the stage by incoming extract; similarly, the normal extract flow from stage $(n+1)$ is accompanied by entrained raffinate and recycle extract carried into stage $(n+1)$ by stage-n raffinate. For perfect stages the entrained phase from any stage will be in equilibrium with the main flow of the other phase in that stage. If it is assumed that oil refractive indexes are additive on a volumetric basis, the following balance can be written around the top of the extractor and stage n:

$$(X+\epsilon_x Y) I_n S_x + \epsilon_y X I'_n S_y = R I_N S_R +$$

$$(Y+\epsilon_y X) I'_{n+1} S_y + \epsilon_x Y I_{n+1} S_x \quad (15)$$

Division of Equation (15) by $(S_y Y)$, S_x and S_y being assumed constant, with the ratio $(S_x X/S_y Y)$ called s and the ratio (S_x/S_y) called R_s, gives

$$sI_n + I_n \epsilon_x R_s + I'_n \epsilon_y \left(\frac{s}{R_s}\right) = sI_N +$$

$$I'_{n+1} I'_{n+1} \epsilon_y \left(\frac{s}{R_s}\right) + I_{n+1} \epsilon_x R_s \quad (16)$$

For a linear equilibrium line with the R.I. coordinates chosen to make this line go through the origin,

$$I'_i = mI_i, \text{ for the } i \text{ th stage.} \quad (17)$$

Substitution of (17) into (16) with rearrangement gives

$$I_{n+1} - A\,I_n = BI_N \quad (18)$$

where

$$A = 1 + \frac{1}{\epsilon_f}\left(\frac{s}{m} - 1\right)$$

$$B = -\frac{s}{m\epsilon_f}$$

$$\epsilon_f = \left(1 + \frac{\epsilon_y s}{R_s} + \frac{\epsilon_x R_s}{m}\right)$$

Equation (18) is a first-order, linear difference equation which can be solved (2) to give

$$I_n = b_1 (A)^n + \frac{BI_N}{1-A} \quad (19)$$

The boundary condition on Equation (19) is that at the feed stage the raffinate-oil refractive index equals that of the feed. Thus

$$I_n = I_F \text{ for n} = 0$$

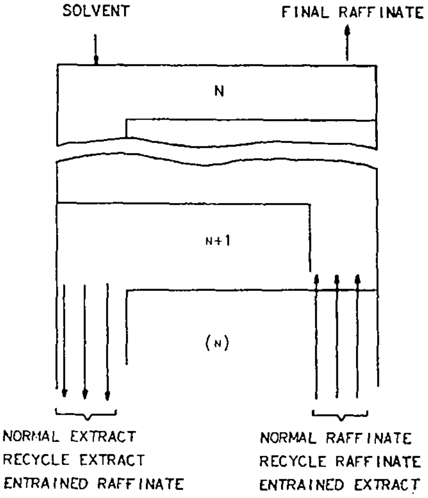

Fig. 5. System for phase entrainment.

Substitution of this condition into Equation (19) gives

$$b_1 = I_F - \left(\frac{BI_N}{1-A}\right)$$

Substituting this value for b_1 into Equation (19) and setting $n = N$, where N is the top stage of the entrained extractor, gives after rearrangement

$$\frac{I_F - I_N}{I_F} = \frac{(1-A-B)(1-A^N)}{(1-A) - B(1-A^N)} \quad (20)$$

The result in Equation (20) can be expressed in terms of only the (s/m) ratio and the entrainment factor ϵ_f by substitution of the definitions for A and B. This substitution gives

$$\frac{I_F - I_N}{I_F} = \frac{1 - \left[1 + \frac{1}{\epsilon_f}\left(\frac{s}{m} - 1\right)\right]^N}{1 - \left(\frac{s}{m}\right)\left[1 + \frac{1}{\epsilon_f}\left(\frac{s}{m} - 1\right)\right]^N} \quad (21)$$

Solving Equation (21) for N with $(I_F - I_N)/I_F = \phi$ gives

$$N = \frac{\log \dfrac{(\phi - 1)}{\left(\dfrac{\phi s}{m} - 1\right)}}{\log\left[1 + \dfrac{1}{\epsilon_f}\left(\dfrac{s}{m} - 1\right)\right]}$$

If no internal entrainment occurs and the extractor operates on the same feed to produce the same final raffinate, the number of theoretical stages required will be

$$N_t = \frac{\log \dfrac{(\phi - 1)}{\left(\dfrac{\phi s}{m} - 1\right)}}{\log\left(\dfrac{s}{m}\right)}$$

as $\epsilon_f = 1.0$.

Hence, if the efficiency of an internally entrained extractor is defined as the ratio of the number of theoretical stages needed for a given feed improvement to those needed when internal entrainment occurs, the efficiency will be

$$\eta_a = N_t/N = \frac{\log\left[1 + \dfrac{1}{\epsilon_f}\left(\dfrac{s}{m} - 1\right)\right]}{\log \dfrac{s}{m}},$$

$$\frac{s}{m} \neq 1.0 \qquad (22)$$

If $s/m = 1.0$, the efficiency follows from Equation (21) by differentiating both numerator and denominator with respect to s/m and evaluating at $s/m = 1.0$. Doing this and solving for efficiency, as above, gives

$$\eta_a = \frac{1}{\epsilon_f}, \frac{s}{m} = 1.0 \qquad (23)$$

where ϵ_f is the entrainment factor. A plot of Equations (22) and (23) is shown in Figure 6.

An estimate of the effect of phase entrainment on the extraction efficiency by use of Equation (22) with typical plant values for the variables is shown in Table 4. For this typical lube extraction the efficiency of a perfect stage decreases rapidly with increase in entrainment in either stream, and simultaneous entrainment in both streams is very serious. The calculation shows also that efficiency is most reduced by entrainment in the extract.

TABLE 4.—EXTRACTION EFFICIENCY REDUCED BY ENTRAINMENT

Operating conditions
Oil in raffinate (S_x), vol. % = 88
Oil in extract (S_y), vol. % = 20
Equilibrium line slope (m), average = 2.5
Operating line slope (s), average = 2.3

A. Entrainment in one phase

Entrainment in	Extract				Raffinate			
Amount entrained, %	2	5	10	20	2	5	10	20
Over-all efficiency, %*	96	91	83	69	99	97	94	88

B. Entrainment in both phases

Entrainment, %, in			
Extract	5	10	5
Raffinate	10	5	15
Over-all efficiency, %*	87	81	84

*Over-all efficiency = apparent efficiency.

The results are only directional for an actual stage which operates at less than 100% efficiency when running entrainment free. It is probable that they do set a limit on the harmful effect of entrainment, as the efficiency of an actual stage would be expected to be less affected by a given amount of entrainment as the unentrained efficiency of the stage decreased. It would appear, therefore, that up to about 5% entrainment in the extract or 10% entrainment in the raffinate would not seriously reduce the efficiency of an actual stage in typical lube extraction service.

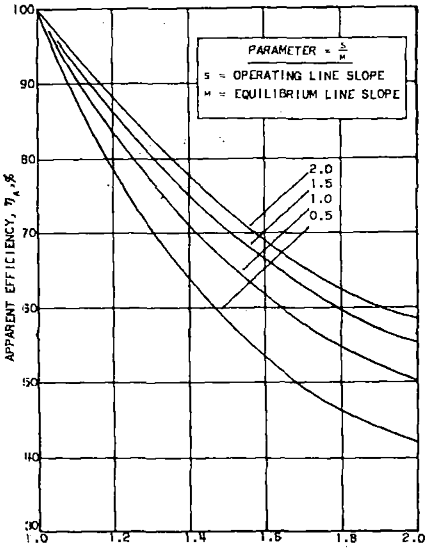

Fig. 6. Effect of internal entrainment on efficiency.

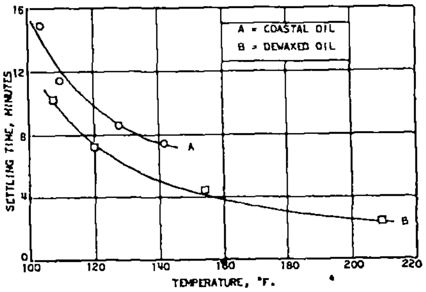

Fig. 7. More rapid separation of phases at higher temperatures.

CAPACITY STUDIES

Maximum stage efficiency, though very desirable, is not the sole criterion of the performance of a commercial extraction tower. Tower capacity, or the maximum flow rates that can be successfully handled without entrainment, must also be considered for a true rating of over-all tower performance. These factors are interrelated, as the presence of entrainment shows that the capacity has been exceeded, and entrainment decreases efficiency. For high capacity the intimate mixture of phases in the mixing section of the stage must separate quickly, and this separation depends on the phsyical properties of the phases and the flow pattern in the settling section of the stage. For a given stage design the physical properties will govern capacity.

Temperature Effect

An arbitrary, though quantitative, test of the settling characteristics of lube-phenol systems was devised to study the effects of phase viscosity and differential density. Equal volumes of phenol and oil were put in a small graduate and mixed with a standard stirrer for a fixed number of strokes while the graduate was maintained at constant temperature. This mixing gave an intimate mixture, or "emulsion," of oil droplets in phenol. The time required for the emulsion to settle into two clear phases was used as a measure of the ease of separation of the phases.

The effect of temperature on the rapidity of phase separation was in-

TABLE 5.—CAPACITY INCREASED BY HIGHER EXTRACTION TEMPERATURES

Type of run	Low temperature	High temperature
Tower top temperature, °F.	180	215
Tower bottom temperature, °F.	160	195
Phenol-water content, vol. %	6.2	8.5
Phenol treat on feed, vol.%	124	124
Final extract-oil content, vol.%	15.8	15.3
Maximum oil rate without entrainment, gal./hr.	0.5 to 0.8	2.0
Settling time, min., in Stage:		
9 (second from top)	6	3
2	8	4

vestigated by getting settling-test data at various temperatures on a wide variety of lube-phenol systems. The oil content of the extract, or phenol phase, was maintained constant at 25 vol. % by varying the phenol water content. Phase viscosity was estimated on the assumption that fluidities were additive on a volume basis. The rapid decrease in settling time associated with increasing temperature, illustrated for two of these stocks in Figure 7, indicates that higher capacity should result with higher extraction temperatures.

The temperature effect was tested experimentally in a 20-stage pilot plant extractor(*1*). For the test work a paraffinic base stock was used which had a gravity of 18.7° A.P.I. and a viscosity of 225 SSU at 100°F. The extractor was operated with a total of ten stages and a final extract oil content of 15%. Extractor feed rate was increased at two temperature levels, while the raffinate quality was maintained, until the entrained oil phase appeared in the final extract. Agitation speed was kept constant in these runs. Results of these tests are shown in Table 5.

With the higher extraction temperature, oil throughput was increased about threefold. As the phenol treat was held constant, the extractor over-all capacity increased by this same amount. It is seen that the increased capacity was accompanied by a decrease in the settling time of the internal tower phases. The results of this study coupled with the effects of temperature on extraction efficiency indicate that extraction temperature is a powerful variable in lube-oil extractions.

Effect of Physical Properties

The settling-test data were generalized by considering that the settling characteristics of the systems could be represented by Stokes' Law. Expressed in terms of the time required to settle a fixed distance, Stokes' Law gives.

$$\frac{1}{\theta} = \frac{g\,(\rho_y - \rho_x)\,D^2}{18\,L\mu_y} \qquad (24)$$

Hence a plot of the reciprocal of settling time vs. the ratio of the differential density to extract-phase viscosity should plot as a straight line. Figure 8 shows the settling-test data plotted in this way; data on kerosene-water were included to take advantage of the low extreme of oil viscosity. The data correlation is quite good with the kerosene-water data falling in line with those for lube-phenol systems. The plot shows that settling time is decreased by an increase in the density difference between phases or a reduction of the extract-phase viscosity. Stages where capacity is critical in a lube extractor will, therefore, be those where the density difference is least and the extract-phase viscosity the greatest. Figure 9 shows the changes in phase densities, differential density, and phase-settling time at flow conditions through a commercial extractor when a paraffin base oil is extracted from a Canadian crude source. The operation was run with the oil feed entering the bottom stage, this stage temperature being about 25°F. lower than the top stage. It is seen that the density differential is smallest and essentially constant in the bottom four or five stages, owing to the extract density reaching a minimum value near the middle of the extractor because of the opposing effects of increasing oil content and decreasing temperature. Since the settling time of the stage phases increases progressively down the extractor, the results indicate that extract-phase viscosity is the governing property in the bottom stages and that the capacity critical stages are the feed stages and those near them.

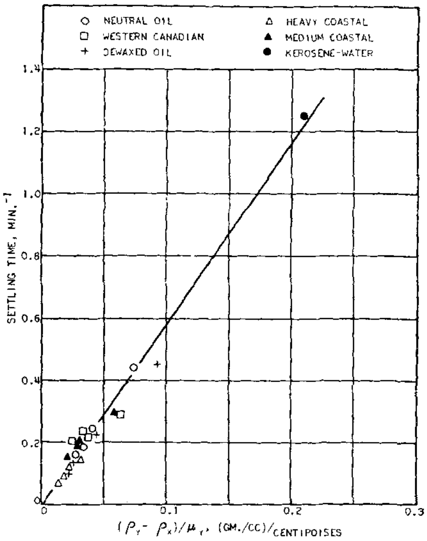

Fig. 8. Lube-phenol settling test following Stokes' Law.

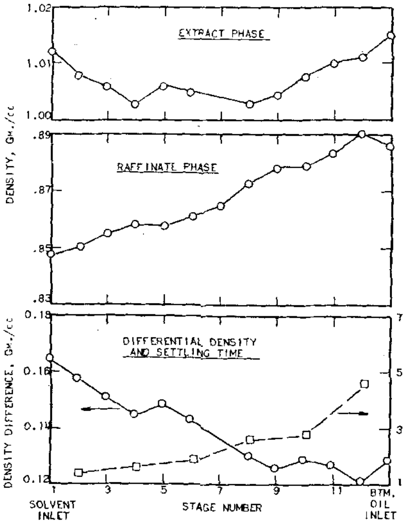

Fig. 9. Phase density and settling time in commercial extractor oil feed: Western Canadian paraffinic lube distillate; viscosity at 100°F. = 78 SSU; gravity = 21.9° A.P.I.

SUMMARY

The principal findings of the extraction studies reported here are as follow.

1. Extraction temperature has a very marked influence on stage efficiency in lube-phenol extractions; higher temperatures give higher efficiencies by reducing oil-phase viscosity. Interfacial tension, though not a major variable in lube extraction, also has a major effect on stage efficiency by affecting drop size.
2. Interstage entrainment in either one or both phases in lube-phenol extraction reduces efficiency markedly, entrainment in the extract being most serious.
3. The flow capacity of extractor towers used in lube service increases with increasing extraction temperature. The capacity limiting stages are the feed stage and the stages near it. Viscosity of the continuous phase in these stages appears to be the governing physical property.

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NOTATION

A = cross-sectional area, L
a = surface area per unit volume, L^{-1}
b_1, b_2, etc. = constants
c = fractional concentration of aromatics in raffinate
c' = fractional concentration of aromatics in extract
D = drop diameter, L
E = extract oil flow, L^3/θ
f = fractional holdup of oil in mixing zone
G = oil flow per unit cross section, L/θ
g = acceleration of gravity, L/θ^2
I = raffinate-oil refractive index
I' = extract-oil refractive index
K = over-all mass transfer coefficient
k = individual film mass transfer coefficient
L = mixing section length or distance, L
m = equilibrium line slope
R = raffinate oil flow, L^3/θ
$R_s = S_x/S_y$ = phase solubility ratio
S = fraction of oil in phase
s = operating line slope in extraction
X = raffinate-phase flow, L^3/θ
Y = extract-phase flow, L^3/θ
ε_x = entrainment of raffinate phase in extract phase, L^3/L^3
ε_y = entrainment of extract phase in raffinate phase, L^3/L^3
ε_f = entrainment factor
η_a = apparent efficiency
η_c = cocurrent efficiency
η_M = Murphree efficiency
η_o = over-all efficiency
μ = absolute viscosity, $M/L\theta$
υ = kinematic viscosity, L^2/θ
ρ = density, M/L^3
θ = time

Subscripts

E = extract
F = feed
R = raffinate
o, x = raffinate phase
y = extract phase

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Thermodynamic Properties of Air

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In recent years the development of new sources and uses for high pressure air has emphasized the need for thermodynamic charts covering a wider pressure range than do the published charts. To fill this need, the chart by Williams (*9*) has been extended to cover pressures from 1 to 1,000 atm. in the temperature range of 300° to 675°R. Previously published temperature-entropy charts for air, all including isoenthalpic lines, are those of

	Temp., °R.	Press., atm.
Gerhart (*2*)	460 to 1,010	1 to 238
N.D.R.C. (*8*)	140 to 760	1 to 238
Williams (*9*)	125 to 500	1 to 220

The charts of both Gerhart (*2*) and Williams (*9*) were calculated from Joule-Thompson expansion data; the values obtained are also given in tabular form. The method for determining the values for the N.D.R.C. (*8*) chart is not reported; this chart is small and difficult to use and tabular results are not given.

The additional values required for the chart presented here were calculated by the thermodynamic techniques outlined below and serve as an independent check on the calculations of both Williams (*9*) and Gerhart (*2*), which differ by less than 0.5% from those given in this paper.

The sources of the data utilized and the range over which each source applied are as follows:

	Temperature, °R.	Pressure, atm.
Amagat (*1*)	460 to 675	100 to 1,000
Holborn and Otto (*4, 5*)	492 to 675	1 to 100
Nelson, Obert "Z" charts (*7*)	300 to 500	220 to 1,000
Williams (*9*)	125 to 500	1 to 220

There are some P.V. data by Koch (*6*) and Witkowski (*10*) in the range 300° to 492°R. and 1 to 220 atm. but they are not of sufficient quantity to be usable. The data which were available, as P.V. data or as Z charts, required calculations and graphical solutions to give the required enthalpy and entropy values (*3*). As a basis for construction of the chart the value of zero entropy has been arbitrarily assigned to saturated liquid air at 1-atm. pressure, and an enthalpy value of 100 B.t.u./lb. has been arbitrarily assigned to gaseous air at zero pressure and 0°R.

The enthalpy of the air at any pressure and temperature is obtained from Equations (1), (2), (3), and (4).

$$H_{PT} = H^*_T - (H^* - H_P)_T \quad (1)$$

$$H^*_T = \int_0^T C^*_p \, dT + 100 \quad (2)$$

$$(H^* - H_P)_T = \int_0^P \frac{RT^2}{P} \left(\frac{\partial Z}{\partial T} \right)_P dP \quad (3)$$

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Note: Figure 1, measuring 17 by 22 in., is available from the authors at a cost of 25 cents for handling and mailing.