

PROCESS DESIGN FEATURES OF A 5 TONNES/DAY MULTI – STAGE, INTERMITTENT DRAINAGE, CONTINUOUS FULL IMMERSION, VEGETABLE OIL SOLVENT EXTRACTION PLANT

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Abstract

In this work, the process design features of a seven stage pilot extraction plant for use in vegetable oil extraction are presented. Three local oils were used in the study. They were Palm oil, Palm Kernel oil and Soy-bean oil. However PKO was used as a case study in the design because it is a simple oil with properties falling between those of palm oil and Soyabean oil. Material and Energy balance calculations were established for the oil as a base of mechanical design of part II. N- hexane was used as a leaching solvent. The efficiency of the plant was found to be 0.70. Process design calculations gave 23.17Kj as the energy requirement of the extractor based on a capacity of approximately 1 Kg of oil extract. The diffusivity of the oils in the solvent averaged $4.0 \times 10^{-9} \text{ m}^2/\text{s}$, while the rate of mass transfer into the solvent was $6.47 \times 10^{-7} \text{ Kmole/m}^2 \cdot \text{s}$, resulting in a mass transfer coefficient of $3.2 \times 10^{-5} \text{ Kmole/m}^2 \cdot \text{s}$.

1. INTRODUCTION

Vegetable oils are compounds of glycerol and other organic acids extracted from plants. They are water insoluble substances consisting of glycerol ester of fatty acid or triglycerides, (Anderson [1] and Hoffmann, [2]. Demand for vegetable oils, either in their refined or raw states, has increased world wide in the past few years due to increasing population, Ige, et al[3], and Van Oss [4], rising standard of living, and consumer preference, Akubugwo, et al [5], Odoemelam, [6] and Kyari, [7]. In the energy sector, vegetable oil will provide the source of bio-energy and renewable energy requirements of the world, Pahl [8] and [4].

Since ancient times, human beings have known how to extract the oils from their various sources and make them fit for their own use. The natives of the tropical regions of the world have long extracted these oils after drying the nuts and seeds under the sun. 4000 years [9]. These oils were consumed raw, since very little or no treatment was made order than filtration and decantation. Mahatta, [10] observed that Olive oil was extracted as far back as 3000BC in ancient Egypt.

Review of related literatures, revealed that much articles have been published on extraction of oils from their mother seeds and nuts using various method, and characterization of the extracted oils. The method by which a particular oil seed is extracted depends on the type of analyzed material(seed), the oil content of the seed and the nature of the subsequent analytical problem, Yeshajahn and Clifton [11], Iwe [12], FAO [13] and Lawson et al, [14]. Harvesting of crop at the right time

is necessary if maximum yield is to be obtained, Vas Oss [4].

Mechanical expression , according to Janet, [15] and Kalu, [16] offer advantage in terms of purity over the solvent extraction, but there is a major draw back when the yield from the method is considered, as yield is almost always barely 65 per cent. According to Othmer and Agrawal [17], mechanical press method is very effective for seeds containing 30 – 70%, oil.

Vidke and Souslski, [18], and Swetman and Head, [19], carried out mechanical expression of oil using different presses; lever and box, hydraulic, Ram, Bridge and Continuous Screw press. Gurnham, and Masson, [20] have studied the equilibrium condition of expression after a constant pressure has been applied and maintained until no further flow of oil occurred. They considered the quantity of oil expressed as the difference between the quantity originally present and that which remained in the cake after expression.. Accordingly, they noted that an increase in pressure on a system of expressible material considered as a fractional increase over the previous pressure causes a proportional increase in the bulk of the solid portion of the system, and hence,

$$\frac{dP}{P} = Kd\rho_S = kd\left(\frac{1}{V}\right) \quad 1$$

$$\text{Or } \frac{dP}{d(1/V)} = kP \quad 2$$

Integration of equation 2 gives

$$\text{Log}P = K + \frac{k'}{V} \quad 3$$

Where K, k and k' are constants, depending on the nature of material, and on the expression conditions, ρ_s is the bulk density of the solid portion of the system, P and V, the pressure and specific volume of the system respectively.

The oil seeds have to be decorticated, cleaned to remove sand, dirt and stones prior to mechanical expression, according to Gurnham and Masson[20]. Kurki [21] observed that chaff in the seeds during expression will absorb some of the oil and keep it from being squeezed out of the expeller.

Waiter and James, [22], opined that solvent extraction is a solid – liquid operation called leaching concerned with extraction of a soluble constituent from a solid by means of a solvent. Perry and Green, [23], called the stream which contains the leached solute the overflow (miscella). [22], had observed that equilibrium is reached when the concentration of oil in the miscella within the pores of the solid is equal to the concentration of oil in the free miscella outside the solid. Coulson and Richardson, [24], concluded that solvent extraction is affected by particle size distribution, temperature, porous structure of the solid, nature of solvent and agitation.

Mass transfer in solvent extraction was also studied by [24], using the concept of thin film as providing the resistance to transfer. They observed that;

$$\frac{dm}{dt} = k'A(C_S - C)/bv \quad 4$$

Where A is area of solid – liquid interphase, b effective thickness of liquid film surrounding the particles, C concentration of solute (oil) in the bulk of the solution (miscella) at time t, C_s concentration of saturated solution in contact with the particle, m mass of solute transferred in time t, and k', diffusion constant. For a batch process in which V, the total volume of the solution is assumed constant,

$$dm = VdC \quad 5$$

$$\frac{dC}{dt} = k'A \left(\frac{C_S - C}{bv} \right) \quad 6$$

The time taken for concentration of the solution to rise from its initial volume C_o to a value C, is found by integration. If it is assumed that A and b remain constant, we have;

$$\int_{C_o}^C \frac{dC}{C_S - C} = \int \frac{k'A}{bv} dt \quad 7$$

$$\text{And } \ln \frac{C_S - C_o}{C_S - C} = \frac{k'A}{vb} t \quad 8$$

Which, for pure solvent, in which $C_o = 0$, becomes

$$1 - \frac{C}{C_S} = e^{-(k'A/vb)t} \quad 9$$

$$C = C_S \left[1 - e^{-k'A/vb)t} \right] \quad 10$$

Equation 10, shows that the solution approaches a saturation condition.

A succinct review of published literatures on liquid-liquid extraction was presented, which clearly explained the important influence of solvent nature on the extent of extraction. Davie and Vincent [25], observed that direct solvent extraction is used for low oil content (< 20% oil), seed such as Soya bean, rice bran and dry milled corn germ. Tray and Bilbe [26], experimented on a homogenous oil impregnated material consisting of thin platelets of uniform thickness with the two phases as the total surface area based on simple diffusion and observed that the theoretical rate of extraction is given by;

$$E + \frac{8}{\pi^2} \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2} e^{-(2n+1)^2 \pi^2} \frac{D\theta}{4R^2} \quad 11$$

Where E is the fractional total oil unextracted at the end of time, θ ; R is one-half of the plate thickness (ft); and D the diffusion coefficient in (ft²/hr). For n = 0, eqn. 11, will reduce to;

$$E = \left(\frac{8}{n^2} e^{-\pi^2} \right) \left(\frac{D\theta}{4R^2} \right) \quad 12$$

Or

$$\log_{10} E = -0.091 - 1.07 \frac{D\theta}{R^2} \quad 13$$

Treybal [27], gave an expression to be used to determine the diffusivity of a liquid in another, considered to be stagnant as;

$$D_A = \frac{(117.3 \times 10^{-18}) \psi (M_B)^{0.5} T}{\mu V_B^{0.6}} \quad 14$$

The rate at which the glyceride of fatty acid diffuses into hexane is given by Bird, [28] as;

$$N_A = \frac{D_{AB}}{z_2 - z_1} \left(\frac{\rho}{M} \right)_{av} (x_{A1} - x_{A2}) \quad 15$$

The mass transfer coefficient is given by Perry and Chilton, [29] as;

$$N_A = K_x(x_{A1} - x_{A2}) \quad 16$$

Fan and co-workers [30], observed two solvent extraction methods, namely:

- Solution extraction, in which oil is extracted from ruptured cells, and
- Diffusion extraction, in which the oil is extracted from unruptured cells.

This gives reason for the variation of diffusion coefficient with time. They noted that prepress, followed by solvent extraction is, however, used for high oil content (>20%oil) seeds such as Palm oil, Palm kernel oil, Cotton seed oil and Groundnut oil. The use of solvent extractors, in our local mills, will therefore, greatly stabilize the processing activities, generate employment and increase the daily throughput.

In the design calculations for an extractor to handle the extraction of the oil from its source, the number of theoretical stages required to ensure effective extraction was calculated from an expression, given by Coulson, and Richardson, [31], as;

$$a^m = \left(\frac{S_o}{S_m} \right) \left(1 - \frac{S_1}{W_1} \right) \quad 17$$

Treybal, [27], also gave an expression from which efficiency can be estimated, namely;

$$E_o = \frac{1}{1 + 3.7(10^4) \frac{KM}{hT\rho}} \quad 18$$

The oil, as extracted, contains a considerable amount of water, as well as some particles, fiber and cell debris, which are responsible for the cloudy nature of the oil, and which must be removed by screening, semi-continuous clarification, or by high speed centrifugation. It also contains carotenoid pigments, phosphotides (gum), odour and oxidation products of Aldehydes and ketones, to the extent of about 0.2%. The average content of free fatty acid (FFA) of vegetable oils is 4%, and hence the need for refining.

In this work, the process design features of an Intermittent Drainage, Continuous Full Immersion, vegetable oil extraction plant is presented, with a view to providing the rural communities of the glob with opportunity of processing their oil seeds and nuts with a cost effective and affordable plant.

2. PROCESS DESIGN

2.1 Process Description

Flaked oil seed (S1) is transferred to the feed hopper (E1) by means of a belt conveyor. Solvent (S2), is also pumped into the same hopper tank. The two components are well mixed to ensure homogeneous mixture, and easy contact of solvent with the flake to leach out the oil. Extraction takes place by the diffusion of the oil into the solvent. The meal (S3), is then transferred to the Launder (E6), for leaching with recycle miscella (S4). The leached meal is then transferred to the next hopper for further extraction of the oil content by means of screw conveyor (E2).

Miscella is transferred through percolation filter (E3), to each of eight (8), miscella sumps. With the miscella transfer pump (E5), the miscella is transferred to the next Launder through valve, (E7), to leach out more oil from the meal. The miscella from the last sump (E8), is pumped to the distillation unit, (E11), for separation, and the distilled solvent (S7), is recycled to the first hopper. The still bottom (S6), is discharged into a decanter (E12), where water (S13), is separated and sent to the sundry water (S14). Steam (S11) is used for distillation and the condensate (S12), goes to the sundry water also. The heat exchanger (E14), is used to heat up the bottom product so as to facilitate separation of water/oil mixture (S9) in the decanter. The cake (S10), with less than 4% oil content is discharged to the desolventizer, (not shown), and the oil extract (S8), is sent to storage. The miscella recycle pump (E9), is now used to recycle the miscella to the first launder, through the control valve, (E10).

2.2 Design Procedure

The procedure assumes that the flow rates, as well as the process steam temperature are known. In addition, the following geometrical data of an extractor are specified: extractor height and diameter, number of stages, hopper and launder dimensions. The physical properties are either specified or are estimated using some correlations and empirical expressions as shown in the report. The steps followed in the design include;

- Calculate the flow rates of flakes and solvent, as well as the miscella
- Use the values obtained in i) above to calculate the number of theoretical stages
- Determine the efficiency of the process
- Use ii) and iii) to determine the number of actual stages desired for the extraction

- v) Determine the heat capacities of the steams to be used in energy balance calculations
- vi) Perform the process design (material and Energy balance) calculations

In performing the material balance of the process, the streams of importance are the solid, liquid, and mixtures of materials that enter and leave each stage in the cascade of extractors.

These include, the prepared palm kernel grits, solvent, extracted oil and solvent mixture, extracted solid material, cooling water and steam.

The generalized law of conservation of mass was expressed as a material balance in order to account for the flow of material into and out of the system, [31], [32], [33], and [34], and for a steady state operation without reaction, we have;

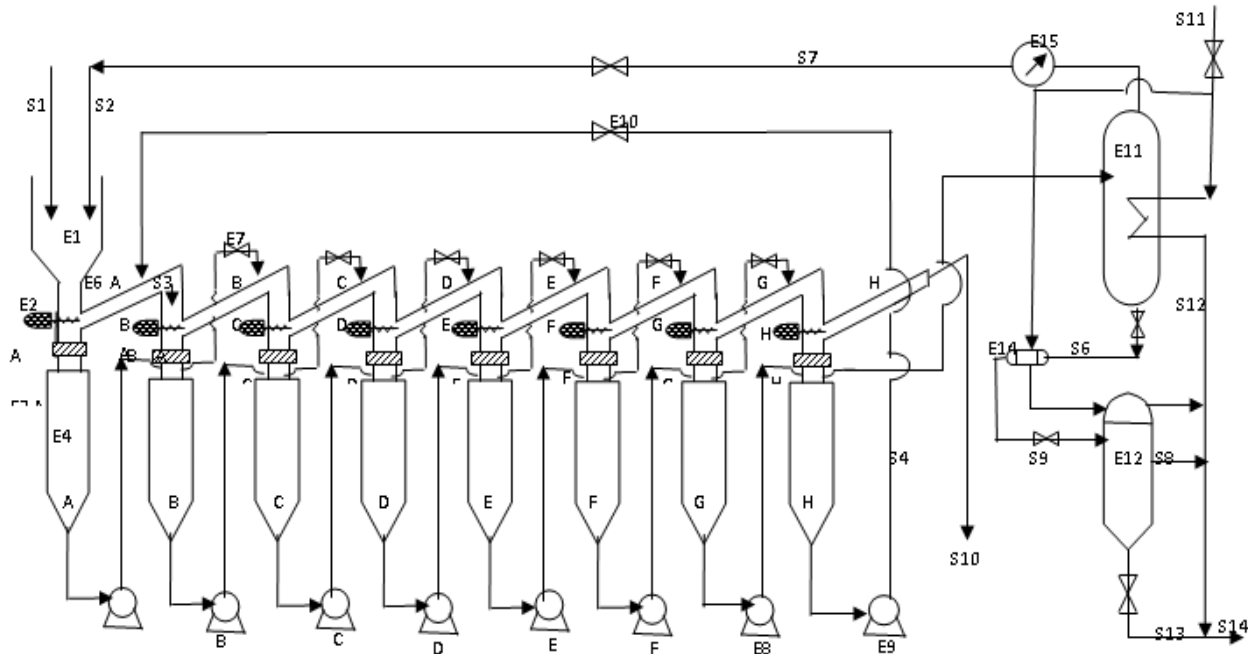


Fig. 1 Flow diagram of a multi – stage intermittent drainage, continuous full immersion, vegetable oil solvent extraction plant.

LEGEND	
E1 – Feed Hopper	S1– Seed flakes to hopper
E2 – Screw conveyor	S2– Solvent
E3 – Percolation filter	S3– Meal (seed cake)
E4 – (A–H) Miscella sump	S4– Miscella recycle
E5 –(A–F) Miscella Transfer Pump	S5– Miscella to still
E6 – Launder	S6– Oil and water mixture
E7 – Valve	S7– Solvent recycle
E8 – Distiller feed pump	S8– Oil to storage
E9 – Miscella recycle pump	S9– Oil/water from H/E to decanter
E10 – Miscella recycle valve	S10– Cake discharge to Desolventizer
E11– Distillation column	S11– Steam
E12– Decanter	S12–Water condenser
E13– Condenser	S13– Waste water
E14– Heat exchanger	S14– Sundry water to treatment
E15– Solvent condenser	

Input through system boundary = Output through system boundary
19

The process design for palm kernel oil extraction was performed for extractor, distillation unit, desolventizer, stripper, and all auxiliary equipment, such as condenser and heat exchange, while, the mechanical design calculations for launder, hopper and sump, and other accessories were done in part II

Design Assumption:

Annual operating time - 300days.

2.3 Data Generation

Palm Kernel Oil – Free Grit Heat Capacity.

Mass of Water = 100g = 0.10Kg
 Initial Temperature of water $T_o = 298K$
 Final Temperature of water $T_f = 373K$
 Heat Capacity of water = 4.187 KJ/Kg. K
 Equilibrium Temperature = 340K
 Mass of Palm Kernel Grit = 0.150Kg
 Initial Temperature of Grit = 298K
 Heat lost by grit = heat gained by water
 $0.1 \times 4.187 (383 - 340) = 0.15 \times C_{pGrit} (340 - 298)$
 $C_{pGrit} (KJ/Kg.K) = \frac{0.100 \times 4.187 \times 43}{0.15 \times 42}$
 $= 2.85 \approx 2.9 KJ/Kg. K$

2.4 Heat Capacity of Palm Kernel Oil

Perry and Green, [23], gave a relationship to be used to obtain the heat capacity of PKO as;

$$C_p (Cal / g. ^\circ C) = \frac{A}{\sqrt{d^{15}}} + B(t - 15) \quad 20$$

Where

t – Reference temperature 25°C (298K) [10]
 d, –Specific gravity of Palm Kernel Oil = 0.9420 **ZZ**
 A and B are constants given by 0.44 and 0.0007, respectively

$$C_p = 0.8869 (Cal / g. ^\circ C) = \frac{3.71KJ}{Kg. ^\circ C}$$

The Heat capacity of Hexane was taken from [23], and shown in table 1

Table 1 Heat capacity of different substances

Streams	Cp KJ/Kg.K
Palm Kernel Grit	2.9
Hexane	2.0
PKO	3.7
Soybean oil	3.6
Soy Flake	0.705
Water	4.187
Steam	4.187

2.5 Number of Stages for Extraction

The results obtained from extraction experiment are;

Weight of PKO used = 1000g
 Amount of oil in grit = 270g

Amount of drained solution before distillation = 1194.2g
 Weight of discharged solid before desolventization = 1123g
 Weight of solvent, oil – free solid = 740g

Weight of drained solution W_m ;
 $W_m =$ Weight of discharged solid with solution - Weight of solvent oil – free solid
 $= 1123 - 740 = 383g$

Percent weight of entrained solution,
 $= \frac{383}{1123} \times 100 = 34.1\%$

Weight of oil obtained after distillation of extract liquid = 260g

Percent weight of oil in miscella
 $= \frac{260}{1194.2} \times 100 = 21.77\%$

Percent weight of oil in Palm Kernel seed
 $= \frac{270}{1000} \times 100 = 27\%$

2.6 Number of Theoretical Stages for Extraction

Definitions:

- m - Number of stages
- S_o - Extracted Oil
- S_1 - Oil in extract solution
- S_m - Residual oil in solid feed
- S_F - Weight of out flowing solute
- W_o - Extracted solution
- W_1 - Weight of miscella
- W_m - Weight of entrained solution
- W_F - Weight of out flowing solution
- a - A characteristic factor for equilibrium stages

Coulson and Richardson, [35], gave the equation for calculating theoretical stages, as;

$$a^m = \left(\frac{S_o}{S_m} \right) \left(1 - \frac{S_1}{W_1} \right) \quad (17)$$

Where;

Moisture content = 8.5%
 Extractable oil = 27%
 Percent oil in solution = 21.77%
 Percent weight of entrained solution = 34.1%

From the data obtained;

On solute; $S_F + S_o = S_1 + S_m$
 On solution; $W_F + W_o = W_1 + W_m$

From the overall material balance, $S_F = 0$
 Basis: 1000g oil – free Grit

$$S_O = \frac{1000 \times 270}{1000 - 260} = \frac{270000}{740} = 364.86g \text{ extracted oil}$$

Similarly; $S_m = 0.1g$ extracted oil residue in solid

Hence,

$$S_1 = S_o - S_m = 364.86 - 0.1g = 364.76 \text{ oil in solution}$$

For 21.77% oil concentration in the solution, weight becomes

$$W_1 = \frac{364.76}{0.2177} = 1675.52g$$

For 1000g grit on oil-free basis, and 34% solution in the solid, entrained solution weight, W_m becomes;

$$W_m = \frac{340}{0.66} = 515.15g$$

But a, is given by, $a = \frac{W_F}{W_m} [35] = \frac{1825.81}{515.15} = 3.54$

Then, from equation 3

$$a^m = \left(\frac{364.84}{0.1} \right) \left(1 - \frac{364.76}{1675.52} \right) = 2854.14$$

Therefore $3.5^m = 2854.14$

$m \log 3.54 = \log 2854$

Number of theoretical stages, m

$$m = \frac{W_F}{W_m} = \frac{\log 2854.14}{\log 3.54} = 6.29 \approx 7$$

2.7 Estimation of Efficiency and Actual Number of

Stages for Extraction:

Efficiency, according to [31], can be estimated from the expression;

$$E_o = \frac{1}{1 + 3.7(10^4) \frac{KM}{hTp}} \quad (18)$$

Where K - Vapour – Liquid equilibrium ratio, y/x; y - Gas

phase concentration at equilibrium, (mole fraction); x - Liquid phase concentration at equilibrium, (mole fraction); M - Molecular mass of the substance, Kg/Kmole; h - Effective liquid depth, mm; ρ - Liquid density, Kg/m³; T - Absolute temperature, K

N-haxane is assumed to be the vapour since it is the more volatile component, (MVC).

The amount of hexane used in the extraction = 0.171Kg
 The molecular mass of hexane is 206Kg/Kmole
 The number of moles of hexane will therefore be equal to = 0.00083Kmoles

Similarly, the number of moles of oil with the molecular mass of 241Kg/Kmole = 0.00042Kmoles

Mole fraction of hexane

$$= \frac{0.00083}{0.00042 + 0.00083} = 0.664$$

Mole fraction of oil = $\frac{0.00042}{0.00042 + 0.00083} = 0.336$

K = $y/x = \frac{0.664}{0.336} = 1.976$

From the extractor cross – sectional area, calculations

$r = 0.18m, D = 0.36m$

If it is assumed that the space above the material on the screw conveyor is 10% of sectional area, then the space that will be created above the conveyor will be;

$$0.1 \times 0.36m = 0.036m$$

Therefore, height of material in the launder

$h = 0.36 - 0.036m = 0.324m$

Temperature T = 333.13K

Now from the calculation of the diffusion,

$$\left(\frac{\rho}{M} \right)_{av} = 4.8459 \text{ Kmole/m}^3$$

$$\left(\frac{M}{\rho} \right)_{av} = 0.2064m^3/\text{Kmole}$$

Then efficiency, $E_o = \frac{1}{1 + 3.7(10^4) \frac{1.976 \times 0.2064}{324 \times 333.15}} = 0.88$

Actual Number of Stages;

Actual number of stages is given by;

$$\frac{\text{Number of theoretical stages}}{\text{Efficiency}} = 21$$

The number of theoretical stages was found to be 7

$$\text{The actual number of stages} = \frac{7}{0.88} = 7.95 \approx 8 \text{ stages}$$

3. SUMMARY OF PROCESS DESIGN CALCULATIONS

3.1 Material Balance

3.1.1 Material balance over Extractor

Basis: 1 second operation

Table 2 Summary of balance over Extractor

Streams	Input (Kg)	Output (Kg)
PK Grit	0.001	
PK Oil		0.00035
Oil in Raffinate		0.000014
PK Inert solid		0.00064
Hexane	0.0017	0.00125
Hexane in Raffinate		0.00045
Total	0.0027Kg	0.0027Kg

3.1.2 Material Balance over Still Column

Basis: 1second operation

Table 3 Summary of balance over Still

Streams	Input (Kg)	Output (Kg)
Miscella	0.0016	
Distillate		0.00125
Hexane		0.0000625
Oil in Bottom W _B		0.00029
Total	0.0016Kg	0.0016Kg

The hexane and the oil form the miscella.

3.1.3 Material Balance over Desolventizer

Basis: 1 second operation

Table 4 Summary of balance over Desolventizer

Streams	Input (Kg)	Output (Kg)
Raffinate	0.0011	
Inert solid		0.00064
Oil in raffinate		0.000014
C ₆ H ₁₄		0.00045
Total	0.0011Kg	0.0011Kg

3.1.4 Material Balance over Stripper

Basis: 1 Second operation

Table5 Summary of balance over Stripper

Streams	Input (Kg)	Output (Kg)
Oil/Hexane feed	0.00029	
Steam	0.0021	
Hexane vapour		0.000064
Water		0.0021
Bottom oil		0.00023
Total	0.0024	0.0024

Table 6 Material balance Process sheet, (Scale up factor = 100.65); Basis: 1 Sec.

Streams	Input (Kg/s)	Scaled value(Kg/s)	up	Output (Kg/s)	Scaled value(Kg/s)	up
Extractor						
PK Grit	0.0010	0.101				
Hexane	0.0017	0.171		0.00125	0.1260	
Miscella	0.0016	0.161		0.00160	0.1610	
PK Oil				0.00035	0.0352	
Raffinate	0.0011	0.111		0.00110	0.1110	
Oil in Raffinate				0.000014	0.0014	
Hexane in Raff				0.000450	0.0453	
Inert Solid				0.000640	0.0644	
Still						
Miscella	0.0016	0.161				
Hexane vapor				0.0012500	0.1260	
Hexane in bottom				0.0000625	0.00625	
Oil in bottom				0.0002900	0.02900	

Desolventizer				
Raffinate	0.0011	0.1110		
Hexane in vapour			0.000450	0.04500
Oil in Raffinate			0.000014	0.00141
Inert in Solid			0.000640	0.06440
Stripper				
Oil-Hexane mixture	0.00029	0.029		
Steam	0.00210	0.211		
Hexane in vapour			0.000064	0.00644
Water			0.002100	0.21100
Oil in bottom			0.000230	0.02320
Total	0.0105Kg	1.0568Kg	0.0105Kg	1.0568Kg

4. ENERGY BALANCE

The scaled up balance sheet of table 6 is used as the basis of energy balance calculation of this section.

4.1 Energy Balance around Extractor

Basis: 1 Second operation

Input and output temperature = 308.15.K and 340.15K, respectively, with 273.15K, as reference temperature.

Table 7 Summary of Energy balance over Extractor

Liquid oil	3.256	
Inert solid	4.669	
Hexane	8.550	8.750
Oil in miscella		4.533
Oil in Raffinate		0.1813
Hexane in Raffinate		3.171
Steam	6.6973	
Total	23.1723	23.1723

4.2 Energy Balance around Still

Basis: 1 Second operation

Feed temperature = 333.15.K, top and bottom temperature = 343.15K and 340.15K, respectively, with 273.15K as reference temperature.

Table 8 Summary of energy balance over Still

Stream	Input (KJ)	Output (KJ)
Miscella	13.283	
Steam	34.775	
Oil		1.073
Hexane		46.86 + 0.125
Total	48.058KJ	48.058KJ

4.3 Energy Balance over Desolventizer

In this column, indirect steam is used to heat up the meal to evaporate the traces of hexane found therein.

Basis: 1 Sec operation

Feed temperature = 333.15.K, top and bottom temperature = 343.15K and 342K, respectively, with 273.15K as reference temperature.

Table 9 Summary of energy balance over Desolventizer

Stream	Input (KJ)	Output (KJ)
Raffinate	9.8893	
Steam	8.1127	
Hexane		16.082
Inert solid		1.920
Residual oil		0.0522
Total	18.002KJ	18.064KJ

4.4 Energy Balance over Stripper

The oil leaving the distillation unit as bottom product contains traces of hexane liquid which must be recovered using direct superheated steam.

Basis: 1 Second operation

Material temperature = 340.15K, Steam temp. = 474.15K, top and bottom temperature = 404.15K, with 273.15K as reference temperature.

Table 10 Summary of Energy balance over stripper

Streams	Input (KJ)	Output (KJ)
Hexane / Oil	.035	2.904
Steam	608.832	
Water		600.38
Oil		6.744
Total	608.867KJ	610.028KJ

Difference in energy of the input from those of the output is 1.131KJ, which is less than 0.2%.

5.0 COMBINED MATERIAL AND ENERGY BALANCE CALCULATIONS

5.1 Introduction:

The hexane from the distillation column and stripper has to be condensed to obtain liquid hexane, which is recycled, while the oil and desolventised inert solid (meal) from the stripper and desolventizer (bottom product) will be cooled by means of heat exchangers using water as cooling medium before discharge.

5.2 Combined Material and Energy Balance over Condenser

Basis: 1 Second operation

Heat Lost by C_6H_{14} = Heat gained by circulating water
 $m C_p \Delta T (v) = m' C_p \Delta T (l)$

Where m - mass of hexane vapour, and m' mass of condensed hexane

$$\text{Now } \Delta H_{C_6H_{14}} = n \int_{343.15}^{298.15} 2 \text{KJ/KgdT} = -11.34 \text{KJ}$$

$$\Delta H_{H_2O} = n \int_{298.15}^{288.15} 4.2 \text{KJ/KgdT} = -42n \text{KJ}$$

$$-42n \text{KJ/Kg} = -1134 \text{KJ}$$

$$n = 2.7 \text{Kg}$$

Therefore mass of cooling water n is 2.7KJ

Energy change in product stream

Total energy Output = Total energy Input

$$\Delta H_{\text{Distillate}} + 11.34 = 46.86$$

$$\Delta H_{\text{Distillate}} = 46.86 - 11.34 = 35.52 \text{KJ}$$

Condensation of Hexane and water vapour

This includes the hexane from the desolventizer and stripper, as well as the water vapour from the stripper

Basis: 1 Sec. operation

Heat lost by entering streams is equal to that gained by cooling water.

Enthalpy change in exit stream:

1. Desolventizer Product Stream

$$\Delta H_{C_6H_{14}(v)} = n \int_{343.15}^{298.15} 2 \text{KJ/KgdT} = -4.077 \text{KJ}$$

2. Stripper Product Steam

$$\Delta H_{H_2O(v)} = n \int_{404.98}^{298.15} 4.2 \text{KJ/KgdT} = -94.85 \text{KJ}$$

$$\Delta H_{C_6H_{14}(v)} = n \int_{343.15}^{298.15} 2 \text{KJ/KgdT} = -1.38 \text{KJ}$$

Total change in exit stream ;

$$-4.077 - 94.85 - 1.38 = -100.307 \text{KJ}$$

Heat lose to the cooling water:

$$\begin{aligned} \text{Total heat Input} &= \text{total heat lose in exit stream} \\ &= 16.092 + 600.38 + 2.904 + (-100.307) \text{KJ} \\ &= 619.376 - 100.307 \text{KJ} = 519.069 \text{KJ} \end{aligned}$$

Mass of water flowing into the condenser

Heat gained by water = Heat lost by condenser

Assumptions:

Let mass flow of water = mKg
 Water inlet temperature = 293.15K
 Water exit temperature = 399.98K
 Now heat gained by water = 519.069KJ,
 Heat equation is $mC_p\Delta T$
 Then, $mC_p\Delta T(\text{cond}) = mC_p\Delta T(\text{water})$
 $= m \times 4.2 (399.98 - 293.15) = 519.069$

$$\text{Mass of cooling water, } m = \frac{519.069}{4.2(399.98 - 293.15)} = 1.16 \text{Kg}$$

5.3 Cooling of oil from stripper

The finished oil product coming from stripper is first cooled to 70°C (347.15K), before being pumped to storage. The cooling of the oil is done by means of cooling water in the heat exchanger.

Basis: 1 Sec operation

Table 11 Summary of combined material and energy balance over condenser

Streams	Input (KJ)	Output (KJ)
Stripper Steam	600.380	
Oil/Hexane	2.904	
H ₂ O/Hexane		100.307
Condenser		519.069
Desolventizer	16.092	
Total	619.376KJ	619.376KJ

Change in Enthalpy of Exit Oil Stream

$$\Delta H_{oil} = n \int_{340.15}^{343.15} 3.7 \text{KJ/KgdT}$$

$$= 0.029 \times 3.7 (343.15 - 340.15) \text{K}$$

$$= -6.634 \text{KJ}$$

Heat on cooling water used

Input of energy = Output of energy
 6.744 = Heat lose to H₂O + 6.634

Heat lose to H₂O = 6.744 - 6.634 = 0.11KJ

Table 12 Summary of combined material and energy balance over heat exchanger

Streams	Input (KJ)	Output (KJ)
Oil from stripper	6.744	
Oil from exchanger		6.634
Water from exchanger		0.110
Total	6.744KJ	6.744KJ

Mass of Water through the Exchanger

Entry temperature = 293.15K
 Exit temperature = 354.98K
 Heat gained by water = 6.634KJ
 Heat gained by water is given by mCpΔT

Then mCpΔT = 6.634
 m x 4.2 (354.98 - 293.15) = 6.634
 m = 0.026Kg

5.4 Energy and mass of water required to cool the desolventizer meal.

The desolventized palm kernel grit is cooled in a heat exchanger before discharge.

Basis: 1 Second operation

$$\Delta H_{oil} = n \int_{340.15}^{334.15} 3.7 \text{KJ/KgdT} = -0.0313 \text{KJ}$$

$$\Delta H_{inert} = n \int_{340.15}^{334.15} 2.9 \text{KJ/KgdT} = -1.12056 \text{KJ}$$

Total exit energy = -0.0313 - 1.12056 = -1.1513KJ

Table 13 Summary of combined material and energy balance over desolventizer meal cooler

Streams	Input (KJ)	Output (KJ)
Energy in heat exchanger	1.8676	
Heat exchanger oil		0.0313
Heat exchanger Inert		1.1206
Water		0.7163
Total	1.8676KJ	1.8676KJ

Let mass of water into the heat exchanger = m
 Input Temperature = 293.15K
 Output temperature = 298.15K

Heat lose to water from heat exchanger = 1.8676 - 1.1206
 = 0.747KJ

Now heat gained by water = mCpΔT = 0.747KJ
 m x 4.2 x 5 = 0.747
 m = 0.0356Kg

6. CONCLUSION

A multi stage, intermittent drainage, continuous full immersion, vegetable oil, solvent extraction plant for use in vegetable oil extraction has been designed has been designed in this work. PKO was used as a case study because it is a simple oil with attributes of those properties of both palm oil and soyabean oil. However the designed extractor is capable of extracting oils from various other oil seeds and nuts. The design is suitable for small scale vegetable oil industries, especially in those sited in the remote and rural areas of the globe, when scaled up. The process design was made on the basis of 1.0Kg of material (PKO). However, with suitable scale-up factor, it can be adapted for a large industrial out-fit.

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