Ultimate Energy Possibilities in Conventional Solvent Extraction

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ABSTRACT

A short historical background and present state-of-the-art for energy consumption in a complete soybean processing plant is provided. This includes bean drying, preparation, extraction, distillation, desolventizing, meal drying and cooling and lecithin production. The paper details how to achieve a much lower energy consumption than is obtained in today's conventional plant operation.

Possible savings in energy are described. These savings are related to the processing of soybeans, but are also valid in some respects for the extraction of presscake from rapeseed or sunflower. All figures given later on are given as kg or as kcal or as kWh per metric ton of processed seed. The figures given relate to soybeans with a moisture content of 12% and finished meal with 12% moisture.

Steam consumption per metric ton of processed soybeans is shown in Table I. The figure of ca. 395 kg was usual until about 30 years ago. Since then some improvements have been made especially in the first distillation step. Steam was replaced by using the hexane/steam vapors exiting from the desolventizer/toaster (DT). Furthermore, the heating of hexane before entering the extractor changed 6-8 years ago from heating by steam to using DT-vapors to heat by direct contact.

The different figures in Table I cannot be regarded as absolute, since throughput capacities, hexane content in extracted flakes and vacuum prevailing in the distillation affect the figures. However, they can be considered as an average for many plants. Some plants that use predesolventizing already have lower steam consumption as shown in the second column.

TABLE I

Steam Consumption per Ton of Processed Seed (kg/ton)

	a	b	c
Seed preparation warming up from 20 C to 60 C	45	45	45
Hexane warming up from 40 C to 55 C	15	–	–
Extractor heating	10	10	–
Desolventizer-toaster (75 C dome-vapor-temperature)	140	140	110
Distillation			
1st stage (25-75%)	60	–	–
2nd stage (75-95%)	10	10	–
Stripper (95-99.9%)	15	15	5
Meal drying according to various systems ca. 40-80	60	60	–
Degumming and lecithin drying	12	12	12
Hexane recovery according to various systems 15-30	23	23	–
Loss by radiation estimates 10-25 according to plant size	15	15	5
Total: kg steam/ton of processed seed	405	330	177

[a]Situation until about 1950: 405.

[b]Present state of art: 330.

[c]Steam consumption possible to achieve: 177.

General state of the art of today per metric ton of processed seed is 300-360 kg of steam, 26-32 kWh, and 1.5-2.5 kg of hexane consumed. The amount of energy used for extractor-warming is valid for a noninsulated extractor. With insulation and sufficiently warmed seed and hexane there is practically no need of extractor-warming.

The second stage of miscella distillation can be operated using flash steam from the plant steam condensate. A rising film evaporator is used and heating is done at atmospheric pressure. The total steam condensate from the plant is fed into this evaporator, and the flash steam resulting is sufficient to raise the miscella concentration from 75% to 95% oil content. Since there is no pressure and no steam trap, the apparatus does not need coding. The condensate flows by gravity into a collecting tank, and the pipe is submerged below water-level. For the start-up or in case of too low oil temperature during shutdowns, a temperature controller opens a steam valve automatically.

For the third stage of the distillation there is equipment now available which was developed only recently. The size of this unit is much smaller than equipment used up to now. It utilizes steam more efficiently, and removal of the remaining 5% of hexane can be done with only 4 kg of steam per ton of seed processed.

The recovery of hexane from vent gases leaving the plant is done now almost exclusively with mineral oil absorption systems. Without calculating the costs for amortization of the equipment, there are costs for steam, electric power and absorption liquid of about US$ 0.20 per ton of processed seed, which, in many cases, is as much as the value of the recovered hexane. There is now equipment to prevent air from entering into the extractor that is entrained with the material to be extracted. If such equipment is installed in a well maintained plant, there are almost no noncondensables to pass through the absorption plant, and as a consequence, there is little hexane to recover during normal operation. The mineral oil system needs only to be in operation during start-up, shut-downs, and during plant stoppage due to equipment failures.

There are excellent possibilities for saving energy in the desolventizing and meal drying systems. Desolventizer/toasters are normally operated at dome-vapor-temperatures of ca. 72-75 C. There are desolventizers for rapeseed and linseed that are operated at temperatures of 80 C and more, as rapeseed and linseed are considered to be difficult to desolventize. There is modern equipment for desolventization now available that can be operated safely at dome-vapor-temperatures of 66-68 C with a degree of desolventization not believed possible up to now.

From Figure 1 can be seen how much steam can be saved by running at the correct lower dome-temperatures. Under the two temperature lines in C and in F, the corresponding steam consumption of the DT per ton of processed seed can be seen in line 1. In line 2, the available kcal of the DT vapors can be found. These vapors are used in the first stage of distillation. Lines A and B give the amount of kcal necessary for miscella concentrating to 75% oil content, plus preheating of hexane for full miscella concentrations

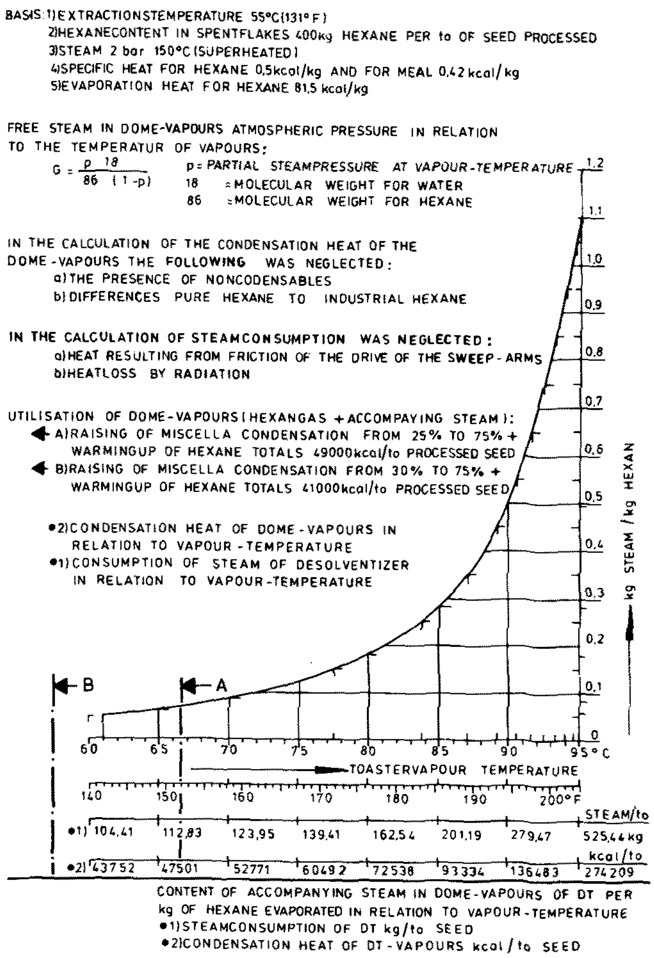

FIG. 1. Influence of dome-vapor-temperature on steam consumption.

of 25% and 30%, respectively. The vertical ordinate shows the steam content in the DT vapors in relation to the vapor temperature.

The diagram shows, for example, the waste of steam for the following condition: dome-vapor-temperature 80 C and full miscella 25%. This shows that ca. 50 kg of steam per ton of seed processed is simply wasted.

The diagram represents a good average for many extraction plants, but cannot be regarded as 100% accurate for any given oil mill. The figures given in Figure 1 are subject especially to hexane content in extracted meal as well as vacuum utilized in the distillation system.

The amount of water to be dried from the meal depends on the amount of direct steam condensing in the meal during desolventizing. With predesolventizing using indirect steam, less steam condenses in the meal. Therefore, only the last traces of hexane are removed with direct steam and in turn less steam is necessary for drying the meal.

It appears that, of all available systems, the most suitable one for predesolventizing is still the application of "Schneckens" or indirectly heated screw conveyors. Mechanical performance is reliable and trouble-free if installed correctly. Their demand for electric energy is relatively low, although thermal efficiency is only 80-85%. Big units have been built, and installation cost pays off in 15-18 months by savings of steam.

A heated screw of 2 m diameter and 7.5 m inside length is good for a heat transmission of ca. 500,000 kcal/hour, corresponding to the evaporation of ca. 6,000 kg of hexane. It is feasible to build units with a capacity of 750,000 kcal.

When ca. 75-85% of the hexane is removed by indirect heat in predesolventizing, then the steam consumption for a DT is reduced to ca. 100 or 110 kg of steam per ton of processed seed. The 110 kg of steam includes also the steam for the predesolventizing. In this way, the moisture content at the exit of the DT is low enough to permit drying in the section of a desolventizer/toaster drier/cooler (DTDC) without the use of steam. One has to be careful not to lower the moisture content during desolventizing too far in order to reach a good toasting effect. The lower limit should be 18% of moisture content at the exit of the DT.

There is another feature to save steam in the drying section of a DTDC. In order to remove the moisture from the meal, there must be sufficient air with a temperature high enough to carry the moisture away. It is possible to heat incoming air to 50 C with the exit air that is carrying the moisture from the drying section of a DTDC by means of a heat exchanger.

Such a heat exchanger is built with a stainless steel shell and glass tubes for the heat exchange. The first such apparatus using this equipment was installed 6 years ago in an oil mill in the Netherlands and today is still operating completely trouble-free.

Inserting the possible savings for desolventizing and meal-drying in Table I gives a total of 177 kg of steam per ton of processed seed.

With reference to the total equipment from extractor to meal drier and cooler in relation to consumption of electric power, nearly all systems such as rotary steam heated driers or fluid-bed-driers and coolers or DTDC-system require a power consumption of about 6-8 kWh if predesolventizing is not used. Using predesolventizing to remove 80-90% of the hexane with indirect heat results in less drying required and the total consumption of electric power drops to ca. 2 or 3 kWh per ton.

Apart from poor maintenance of stuffing boxes etc., hexane losses of a plant occur mostly with high solvent content in meal and vent gases leaving the plant. Hexane contents in meal at the exit of the DT differ with different meals processed, from less than 0.05% by weight in case of soy and up to 0.15% and more in case of rapeseed and linseed meal. With a recently modified DT which has been in operation for 5 months, these figures have been reduced by about two-thirds.

The loss of hexane in an absorption system depends directly on the quantity of noncondensables in the system. This can be brought to almost zero by the application of an extractor feeding device which prevents air from entering into the extractor, so that hexane losses are considerably less than 1 kg.

The costs for energy differ greatly from country to country. For the calculations shown in Table II, the figures chosen are a bit on the high side, since energy prices will increase further in the future, even in countries with relatively low costs at present. It probably surprises you how much energy and money still can be saved for plants processing 300,000 or 500,000 or even a million tons per year.

Costs can be lowered by cogeneration of electricity. Several oil mills have such equipment. Cogeneration depends much on negotiations with producers of electric power, and in many cases it is difficult to reach agreements.

There is also the possibility of utilizing the warm water from condensers for heating purposes or upgrading it by means of heat pumps. In case of using air-operated con-

TABLE II

Cost of Energy in Various Countries and Savings Possible to Achieve

Steam per ton from US $ 12.-20.00 for calculation			US $ 18.00
kWh from US $ 0.04-0.08 for calculation			US $ 0.07
kg of hexane from US $ 0.20-0.45 for calculation			US $ 0.40
Current consumption	330 kg steam	29 kWh	2 kg hexane
Consumption possible to achieve	170 kg steam	25 kWh	1 kg hexane
Difference of consumptions	160 kg steam	4 kWh	1 kg hexane
Multiplied by cost US $	2.88	0.28	0.40
Total savings possible	2.88+0.28+0.40 = US $ 3.56/ton of seed processed.		

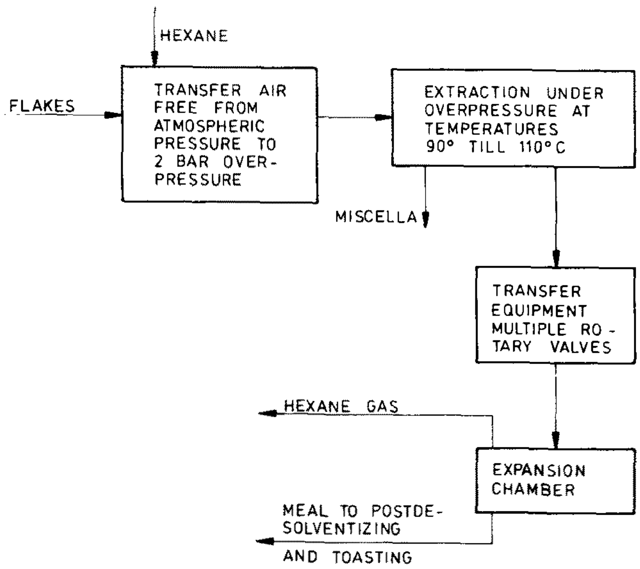

FIG. 2. More ideal extraction system.

densers, the warm air can be utilized for several purposes: for example, it could be used to warm up cracked beans before flaking. In this way, use of 45 kg to warm the seed could be reduced to ca. 15 kg.

There should still be an outlook for future developments and I suggest a system such as shown in Figure 2. In this scheme, the main part is an extraction system under pressure of ca. 1.5-2.5 bar overpressure and at temperatures around 95-115 C.

To transfer the flakes to be extracted from atmospheric pressure continuously in and out of a chamber of higher pressure is technically possible. The latent heat of the solids plus hexane contained at 115 C to be flash-cooled to ca. 70 C is enough to desolventize ca. 85-90% of the hexane in the flakes. The degree of flash desolventizing depends on the pressure and temperature applied in the extractor. Commonly used extractors of today, especially these designed for high capacities, can hardly be reinforced to withstand this pressure. However, there are certain types that can be easily designed at very low cost and be built strong enough to resist the pressures mentioned.

Using the system of extraction under pressure gives many advantages: very short extraction time, and as a result of this, lower investment costs for the extractor, predesolventizer, postdesolventizer/toaster and dryer equipment; no enzyme activity during extraction should result in lower residual lecithin content in degummed oil; and still lower consumption of steam and power.