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RECOVERY OF WASTE ENERGY IN THE PRODUCTION OF EXPANDED SHALE AGGREGATE

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Mineral Processing Division

January 1974

Technical Bulletin TB 184

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Information Canada
Ottawa, 1974

Mines Branch Technical Bulletin TB 184

RECOVERY OF WASTE ENERGY IN THE PRODUCTION OF EXPANDED SHALE AGGREGATE

by

V.V. Mirkovich*

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SUMMARY

In view of increasing fuel costs and the need for energy conservation, a typical process for expanding shales to light-weight aggregate was studied. It was decided that, for such a granular material, counterflow, moving-bed heat exchangers should be considered for recovery of heat from the exhaust gases and from the product. The depths of the beds were calculated after the method of Munro and Amundson. An overall heat balance of the process indicates that 36 per cent of the heat presently used can be saved.

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Bulletin Technique de la Direction des Mines TB 184

RECUPERATION DE L'ENERGIE PERDUE
DANS LA PRODUCTION D'AGGREGATS SCHISTEUX EXPANSES

par

V. V. Mirkovich*

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RESUME

En considération du coût croissant des combustibles ainsi que du besoin de conservation de l'énergie, un procédé typique servant à l'expansion des schistes dans la production d'aggrégats légers a été étudié. Il fut décidé que pour une telle matière granulaire, des échangeurs à température de type contre-courant et à lit mobile devraient être considérés pour la récupération de la chaleur des gaz d'échappement et du produit. Les profondeurs des lits ont été calculées d'après la méthode de Munro et d'Amundson. Un bilan global de la chaleur du procédé indique que 36 pour cent de la chaleur présentement utilisée peut être épargnée.

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CONTENTS

INTRODUCTION

In view of the increasing cost of energy in the processing of minerals and ores, a project was initiated to examine, from theoretical and engineering points of view, the feasibility of increasing thermal efficiency of the production of lightweight concrete aggregate.

As a basis for the following analysis, a typical industrial calcination of shale is described. A shale from the Dundas-Meaford formation, 5/16 to 1 1/4 inches in size, is processed in a rotary kiln, 8 feet external diameter and 135 feet long. The kiln is lined with a 6-inch layer of refractory bricks. The material is fed at the rate of 10 tons per hour, with a retention time of 55 minutes. Heat is supplied by a natural-gas burner at the rate of 1,500,000 Btu/hr ton. The temperature of the material increases from ambient at the feed end to about 2050°F in a hot zone of the kiln. It is discharged at about 1850°F. There is no waste-heat recovery from the exhaust gases; however, an undetermined amount of heat is recovered in the finished-product cooling device at the discharge end of the kiln. The exhaust gases, which leave the kiln at 1050°F, cannot be cooled to temperatures below 500°F because of condensation problems.

In this analysis of a proposed process a vertically moving bed of packed solids is considered as the medium for achieving heat recovery from exhaust gases and finished product. The analysis is based on the described process. An overall heat balance is made to estimate the maximum quantity of recoverable waste heat.

DATA AND CALCULATIONS

Although substantial information on heat-transfer data can be found in the literature, these data are often of a general nature and are not always useful in dealing with specific materials. Therefore, to obtain values for these specific cases, one has to resort either to complicated and time-consuming measurements, or to varying degrees of estimation.

A. Recovery of heat from exhaust gases

The calculations are based on a counterflow heat exchanger consisting of a vertical cylindrical container in which solid particles are introduced at the top and move uniformly downward while gas flows upward through the voids. The constant cross section of the bed is assumed to be 10 sq ft. As the total feed to the kiln is 10 ton/hr, the resulting superficial mass velocity of the feed is 2000 lb/hr ft^2. The particles are considered to be 1 inch in diameter. The heating fuel is natural gas consumed at the rate of 1500 ft^3/hr ton (1,500,000 Btu) by burning with 100% excess air according to reaction,

$$CH_4 + 4O_2 + 16N_2 \longrightarrow CO_2 + 2H_2O + 2O_2 + 16N_2$$

Thus for every 1500 ft^3 (66.8 lb) of fuel gas there will be 2473 lb of combustion products. In addition to this quantity there will be some generation of gases resulting from LOI and moisture in the material. Therefore, the total amount of exhaust gases will be about 2600 lb/hr ft^2 ton.

Experimental and theoretical studies of heat transfer have been made by several authors. Furnas[1, 2, 3] considered heat transfer in a moving bed in an approximate manner. Lowell and Karnofsky[4] examined the problem with the intention of developing an exact method and solved it by a modification of the Schmidt graphical method. An analytical solution for spherical particles was obtained by Munro and Amundson[5] but because answers can be obtained considerably faster by numerical methods utilizing computers than by the rather cumbersome graphical system, the Munro-Amundson method was selected for these calculations. Their method takes account of the thermal diffusivity of the solids and makes allowance for the thermal gradients in the solid. The surface heat transfer coefficient is assumed to be constant. Their solution involves the relationships among four dimensionless ratios, on the basis of which curves (Figure 1) can be calculated for various conditions of heat transfer by using their equations 18 and 22 or 21' and 22'. These ratios are:

$$\varepsilon = \frac{k_s}{hR}, \quad \beta = \frac{G_s c_s}{G_o c_p}, \quad X = \frac{\gamma \chi}{R^2}, \quad Y = \frac{T - t_o}{T_o - t_o}$$

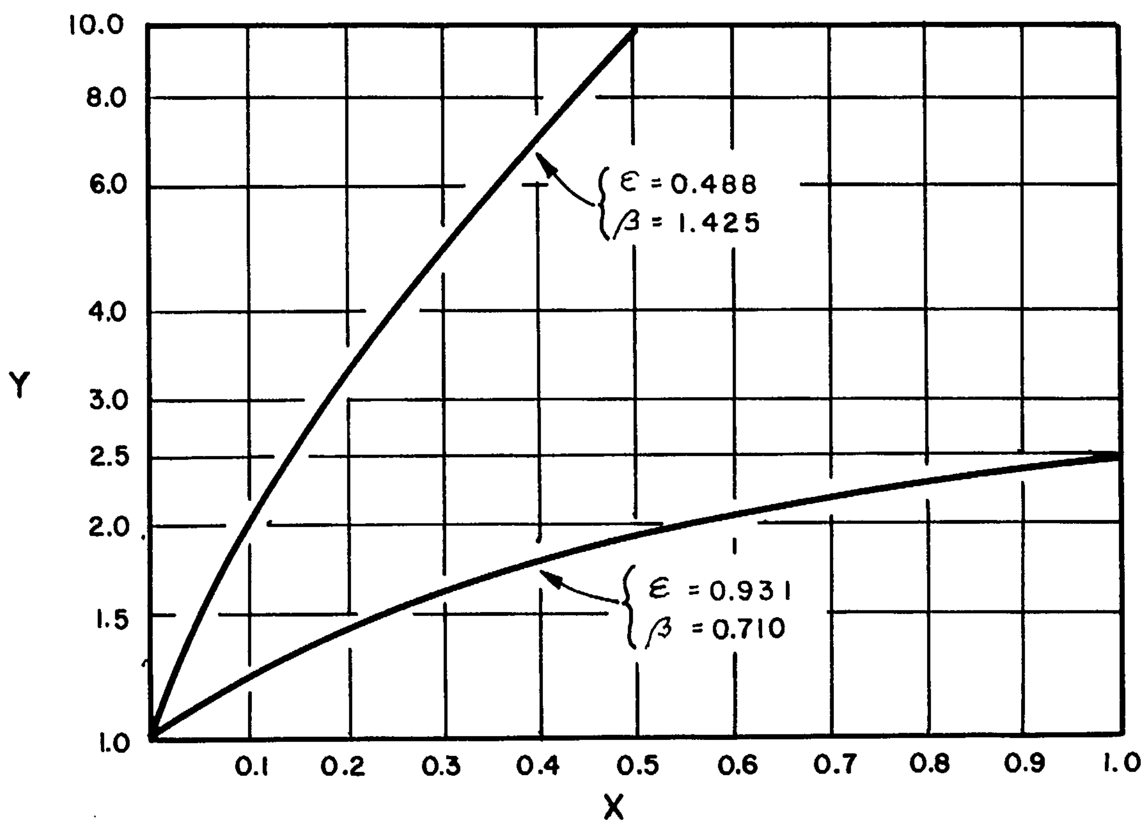

Figure 1. Plot of dimensionless factors X versus Y

In the case under consideration, the values assigned to the different symbols are as follows:

Symbol and Description	Value	Reference
c_p (specific heat of fluid, 650°F)	0.26 Btu/lb °F	McAdams[6]
c_s (specific heat of solids)	0.24 Btu/lb °F	estimated
D (particle diameter, 1 inch)	0.0833 ft	
f(void fraction)	0.5	estimated
G_o (superficial mass velocity of fluid)	2600 lb/hr ft^2	
G_s (superficial mass velocity of solid)	2000 lb/hr ft^2	
h (coefficient of heat transfer between fluid and solid, 650°F):		

$$\frac{DG_o}{\mu} = \frac{0.0833 \times 2600}{0.0725} = 2.99 \times 10^3, \text{ then}$$

from figure 11-10 in McAdams[7]

$$j = 0.036 = \frac{h}{c_p G_o} \left[\frac{c_p \mu}{k_o}\right]^{2/3} = \frac{h}{0.26 \times 2600} \sqrt[3]{\left(\frac{0.26 \times 0.0725}{0.027}\right)^2}$$

$$h = \ldots\ldots\ldots\ldots\ldots\ldots 30.9 \text{ Btu/hr °F ft}^2$$

k_o (thermal conductivity of fluid, average 650°F)	0.027 Btu/hr ft°F	McAdams[6]
k_s (thermal conductivity of solid)	1.2 Btu/hr ft°F	estimated
R (particle radius)	0.0417 ft	
T (gas temperature entering heat exchanger)	1050°F	
T_o (gas temperature leaving heat exchanger)	500°F	
t_o (initial temperature of solid)	50°F	
χ (depth of heat exchanger bed, ft)		
$\gamma = k_s(1 - f)/G_s c_s$		
μ (viscosity of fluid, 650°F)	0.0725 lb/hr ft	McAdams[8]
[ρ (density of solid)	162 lb/ft^3	estimated]

Substituting the above values into dimensionless ratios ε, β and Y, and using the appropriate curve in Fig. 1, the depth of the bed can be established. Thus,

$$\varepsilon = k_s/hR = 1.2/30.9 \times 0.0417 = 0.931$$
$$\beta = G_s c_s/G_o c_p = 2000 \times 0.24/2600 \times 0.26 = 0.710$$
$$Y = (T - t_o)/(T_o - t_o) = (1050 - 50)/(500 - 50) = 2.2$$

From Fig. 1 and curve $\varepsilon = 0.931$ and $\beta = 0.710$, when $Y = 2.2$, one finds that $X = 0.72$.

Therefore

$$X = \gamma\chi/R^2 = 0.72$$

where $\gamma = k_s(1 - f)/G_s c_s = 1.2(0.5)/2000 \times 0.24 = 0.00125$

and $R^2 = 0.001736$

The depth of the heat exchanger bed, $\chi = 1.0$ ft

The quantity of heat transferred from gas to solids (i.e., recovered heat) is:

$$G_o c_p (T - T_o) = 2600 \times 0.26 \times 550 = 371{,}800 \text{ Btu/hr},$$

which is equivalent to: (371,800/1,500,000) 100 = 24.8% of the input into the kiln.

Average temperature of the material leaving the heat exchanger is:

$$[G_o c_p(T - T_o)/G_s c_s] + 50 = [371{,}800/2000 \times 0.24] + 50 = 824°F$$

It should also be noted that there will be a certain pressure drop through the bed of material. Norton(9) gives values for air as 7.7 and 6.9 inches of H_2O per foot of bed height for 5/16 inch and 1/2 inch pebble beds respectively.

B. Recovery of Heat from the Product

In sizing the heat exchanger for the finished product, the change of the physical properties caused by the expansion of the material should be taken into consideration. The critical property in this case is the density. Referring once again to Norton(9), it appears that a gas velocity of 2400 to 2600 lb/hr ft^2 would be too high because it might lift the expanded, lighter material. Consequently, calculations will be based on a moving bed of 19.6 sq ft (5 feet in diameter) cross sectional area. The data which differ from the ones given in the previous chapter are listed below:

Symbol and Description	Value	Reference
D (particle diameter, 118 inches)	0.15 ft	estimated
f (void fraction)	0.5	estimated
G_o (superficial mass velocity of air)	1250 lb/hr ft^2	
G_s (superficial mass velocity of solid: due to ignition and water loss the weight of the original 2000 lb of raw material is reduced to 1900 lb)	950 lb/hr ft^2	estimated
h (coefficient of heat transfer between air and solid)	16.4	
k_o (thermal conductivity of air, 800°F)	0.03 Btu/hr ft°F	McAdams(6)
k_s (thermal conductivity of solid)	0.6 Btu/hr ft°F	estimated
R (particle radius)	0.075 ft	
T (temperature of hot material)	1850°F	
T_o (temperature of cooled material)	300°F	
t_o (temperature of cold air)	50°F	
μ (viscosity of air, 800°F)	0.08 lb/hr ft	McAdams(8)

The depth of the moving bed can now be established by using Munro-Amundson's method. Substituting the above values into the dimensionless ratios, one obtains:

$$\varepsilon = 0.6/(16.4 \times 0.075) = 0.488$$

In this case the solid material is the hotter substance and the role of temperatures (previously the gas was at higher temperature) is reversed. To compensate for this the reciprocal of β must be used, i.e. $\beta = G_o c_p / G_s c_s$. Therefore,

$$\beta = (1250 \times 0.26)/(950 \times 0.24) = 1.425$$

$$Y = (1850 - 50)/(300 - 50) = 7.2$$

From curve $\varepsilon = 0.488$ and $\beta = 1.425$ in Fig. 1, for a value of $Y = 7.2$ results in $X = 0.401$. Therefore,

$$X = \gamma\chi/R^2 = 0.401,$$

where $\gamma = k_s\ (1 - f)/G_s c_s = 0.5 \times 0.6/950 \times 0.2 = 0.00132$

and $R^2 = 0.005625$

and the depth of the bed is:

$$\chi = 1.7 \text{ ft},$$

The quantity of heat recovered from the finished product is:

$$G_s c_s\ (T - T_o) = 1800 \times 0.24 \times (1850 - 300) = 669{,}600 \text{ Btu/hr}$$

Temperature of the preheated air is:

$$G_s c_s \ (T - T_o)/G_o c_p = 669{,}600/(2600 \times 0.26) + 50 = 1041°F$$

HEAT BALANCE

The heat balance is calculated for the present method of production, based on 1 ton of shale per hour. It establishes the requirement for heating (1) raw material from ambient temperature (50°F) to the temperature at the discharge end of the kiln (1850°F), (2) combustion air from ambient temperature to the temperature at the exhaust end of the kiln (1050°F), and (3) the heat losses from the shell of the kiln.

Solids:	$G_s c_s \Delta T$ = 2000 x 0.24 x (1850 - 50) = ...	864,000 Btu/hr
	Decomposition of $CaCO_3$, $MgCO_3$ and other endothermic reactions (estimated)	10,000
	7% moisture:	
	(a) Latent heat: (lb H_2O) x (lat heat of vap) = 140 x 970 = ...	135,800
	(b) Specific heat: (lb H_2O) x (sp heat H_2O)ΔT = 140 x 0.5 x (1050* - 50) = ...	70,000
Air (combustion and excess):	$G_o c_p \Delta T$ = 2600 x 0.26 (1050 - 50) = ...	676,000
Heat requirement without kiln-shell losses:		1,755,000 Btu/hr

Kiln-shell losses:

Approximate temperature distribution on the kiln surface is:

(1) Hot zone: length: 30 ft
average temp : 500°F
(2) Intermediate: length: 40 ft
av temp : 250°F
(3) Preheat zone: length: 65 ft
av temp.: 150°F

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*1050°F is the temperature at which water vapour leaves the kiln together with combustion gases.

Heilman[10] and MacMillan[11] give values for heat-transfer coefficients for simultaneous heat loss by convection and radiation ($h_c + h_r$) from oxidized steel pipes. Values applicable in this case are as follows:

(1) Hot zone $(h_c + h_r)_1 = 4.5$ Btu/hr°F ft^2
(2) Intermediate zone $(h_c + h_r)_2 = 2.7$ Btu/hr °F ft^2
(3) Preheat zone $(h_c + h_r)_3 = 2.15$ Btu/hr °F ft^2.

Heat loss from any section of the kiln:

$(h_c + h_r)$ x (diameter of the kiln) x (length) x Π x ΔT

therefore,

(1) $4.5 \times 8 \times \Pi \times 30 \times (500 - 50) =$	1,526,814 Btu/hr
(2) $2.7 \times 8 \times \Pi \times 40 \times (250 - 50) =$	542,867
(3) $2.15 \times 8 \times \Pi \times 65 \times (150 - 50) =$	351,230
Total kiln losses	2,420,911 Btu/hr

Heat losses from the shell per ton of material: 242,100 Btu/hr ton

Total heat requirement for processing raw material

$$1{,}755{,}000 + 242{,}100 \simeq 1{,}997{,}000 \text{ Btu/hr ton.}$$

It follows that the difference between the total heat requirement (1,997,000 Btu/hr ton) and the heat input to the kiln (1,500,000 Btu/hr ton), which amounts to 497,000 Btu/hr ton, is presently being recovered in the material cooler. Installation of counter-current heat exchangers for preheating combustion air and solid material, without increasing the feed input or decreasing the air supply, would realize the following saving in fuel:

Heat recovered in solids preheater:	371,800 Btu/hr ton
Heat recovered in air preheater:	669,600
Total	1,041,400 Btu/hr ton
Minus the heat presently recovered in the finished material cooler	497,000 Btu/hr ton
Maximum recoverable heat in the system	544,400 Btu/hr ton

which, on the basis of the above consumption, represents a saving of 36%.

DISCUSSION AND CONCLUSIONS

Presumably the first question that arises after examining the calculations in the previous chapters is the problem of the reliability of the basic data. A direct answer to this question is not possible. However, the results should give an indication of confidence that one can have in the estimated data. For example, by using counterflow heat exchangers with the present kiln, the maximum waste heat recovery is not likely to be much higher than the calculated 36%, simply because the heat losses are due to losses in the exhaust fumes (where the temperature cannot be lowered below 500°F), endothermic reactions (such as the evaporation of moisture in the material), and radiation from the surface of the kiln. The value of the first two items can be fairly accurately and readily established. However, the heat losses from the surface of the kiln is another matter. Not only because the surface temperatures are difficult to establish but because the heat transfer coefficient ($h_c + h_r$) can vary considerably under different conditions of draught and wind. Thus, the overall heat requirement could be higher than calculated but, on the other hand, one can see that even if the surface heat losses were 50% higher than estimated (i.e., an increase of 121,000 Btu above the calculated 242,000 Btu), the overall heat requirement would increase not more than 7 to 8%. This means that the heat exchangers would still be able to recover about 30% of the heat presently wasted.

The depths of 1.0 foot for the exhaust-gas heat exchanger and 1.7 feet for the finished product exchanger are probably somewhat overproportioned. Calculations and data were based on the assumption that the solids are regular spheres. For instance, Lowell and Karmofski[(4)], in solving a problem similar to the one considered in this analysis, take the heat-transfer coefficient between exhaust gases and limestone (in a vertical lime kiln) to be 78 Btu/hr °F ft^2!

Particle size has considerable effect on the designed depth of the bed. By changing the size of the particle one affects two heat transfer parameters: a larger particle has lower surface to volume ratio, consequently (1) less heat transfer surface is available per unit weight, and (2) the path of heat travel is longer, therefore requiring longer heat exposure to achieve the same temperature level as in a smaller particle. For example, the depth of the raw material preheater bed would have to be increased from 1.0 to 1.5 ft if the particle diameter is increased from 1 inch to 1.5 inches.

The information given in this report should serve as a base for sizing heat exchangers for recovery of waste heat from existing installations. However, to attain the optimum design of these heat exchangers, it would be advantageous to obtain additional information. Experimental trials on a laboratory scale should be made to establish the maximum superficial mass velocity of fluid (G_0) (conversely, the minimum diameters of the heat exchangers) for different particle sizes of raw material and finished product. Also, a literature survey and actual investigation of operation of existing counter flow heat exchangers should be undertaken. Perry(12), for example, quotes heat recoveries in some European installations to be as high as 80%.

If the construction of the existing material cooler permits, an increase of heat recovery could be realized by increasing the depth of the bed on the cooler and by decreasing the gas flow through the bed. Although the temperature of the air would be lower than in counterflow heat exchanger, if the total flow were reduced to only 100% excess air and all of it is directed into the kiln, the amount of heat recovered would be appreciably increased.

ACKNOWLEDGEMENTS

The author wishes to express his gratitude to Mr. H.S. Wilson for his generous contribution of time in discussing the problem and for many helpful suggestions. Also, thanks are due to Dr. J.A. Soles, who graciously volunteered to prepare a computer program for solution of a transcendental function.

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