REFERENCES

1. Robert.H.Perry and Cecil.H.Chilton. Chemical Engineers Handbook 5th ed. New York Mc. Graw Hill Book Co., 1973.

2. Clyde Orr, Jr. Particulate Technology. New York The Macmillan Company, 1966.

3. Nicholas Pintauro. "Agglomeration Process in Food Manufacture Noyes Data Corporation, New Jersey, 1972.

4. R.G.Gidlow and A.D. Mills."Pillsbury Fluidized Bed Process" U.S. Patent 2,995,773 ; August 15, 1961.

5. Leon I Maissel & Reinhard Glang. Handbook of Thin Film Technology Mc. Graw Hill Inc., 1970.

6. A.M. Swanson and D.J. Fenske. "Lactose Agent for Dried Egg" U.S. Patent 3,262,788 ; July 26, 1966.

7. F.E. Reimers, M.D. Miller and E. Naborney. "All Purpose Sugar Process" U.S. Patent 3,143,428 ; August 4, 1964.

8. J.O. Hardesty, "Agglomeration-A Chemical Engineering Tool for Granulating Mixed Fertilizers" Chem. Eng. Progr., 51 p.291 295, 1955.

9. R.K. Mcqeary."Mechanical Packing of Spherical Particles" J.Am. Ceram. Soc, 44, p. 513-522, 1961.

10. E.J. Crosby, and W.R. Marshall, Jr. "Effects of Drying Conditions on the Properties of Spray-Dried Particles" Chem.Eng. Progr. Vol. 54, No7, p.56-63, July 1958.

11. Ryong-Joon Roe. "Theory of the interface between polymers or polymer solutions." J.Chem. Phys. Vol. 62, No.2p 490-9 1975.

12. akuhei Nose "Theory of Liquid-Liquid Interface of Polymer Systems" Polymer Journal, Vol.8, No.1p.96-113, 1975.

13. Daizo Kunii and Octave Levenspiel. Fluidization Engineering John Wiley and Sone Inc., 1969.

14. Earle A. Clifford. "A Practicle guide to LP-Gas Utilization" L.P. Gas Handbook Series Volume 1, 1960.

15. Colin Carmichael. "Design and Production Volume" Kent's Mechanical Engineers Handbook 12[th] edition 1971

16. John Castell-Evans."Physico-Chemical Tables Volume IIV Physical and Analytical Chemistry, Charles Griffin & Co. Ltd., 1911.

17. Alan Charles Sturt "Polymerization Process."British Patent 1,351 624 May 1974

18. William John Megowen"Method of Agglomerating." U.S. Patent 3,447,962, June 3, 1969.

19. E.J. Crosby and W.R. Marshall Jr."Effect of Drying Conditions on the Properties of Spray-Dried Particles."Ch.E. Prog., Vol.54 No.7, p.56-63, July 1958.

20. F.Frief."Berkeley Physico Course Volume 5." Statistical Physics Mc Graw Hill Book Co., p.44-45, 319-342, 1965.

21. Bird, Stewart & Lightfoot. Transport Phenomena, Wiley International Edition, Wiley &Sons Inc., 1960.

22. ข่าวกรมวิทยาศาสตร์ ฉบับที่ 84, พฤษภาคม 2520, หน้า 3, "พิษตะกั่วจากน้ำมันเบนซิน"

23. Chemical Abstract "Content of Six-membered Naphthene in Gasoline" Volume 68, No.51619 V, 1968.

24. Robert Noyes, "Dehydration Process for Convenience Food" Food Processing Review No.2 Noyes Development Corporation, London, 1969.

25. A.S. Forst. et.al. Principles of Unit Operations, Lehigh University Pennsylvania, 1960.

26. Warren L. McCabe and Julian C., Smith Unit Operations of Chemical Engineering, 2nd ed.Mc.GrawHill Chem. Engin meering Series 1967.

27. Chemical Abstract "Hydrocarbon in 117-23° Fraction of Gasoline" Vol.68, No.51618 U, 1968.

28. Chemical Abstract "Napthene for Producing Gasoline" Vol. 75,No. 38626 V, 1971.

29. Chemical Abstract "Olefins in Gasoline" Vol.87, No.55437,1977.

30. Chemical Abstract "Chemical Composition of Gasoline." Vol.87, No.55438 a, 1977.

31. Chemical Abstract "Gasoline Production." Vol.87, No.55436 y,1977 .

32. The British Plastics Federation"Polystyrene Materials." A code of Practice Copyright Publication,Vol.47,1963.

33. Henry I.Bolker. Natural Synthetic Polymers, An Introduction. Marcel Dekker Inc, Ny. 1974.

34. Esso Standard Thailand Ltd. "Typical Inspection of Esso Gasoline" July 1974.

A P P E N D I C E S

Appendix A

A.1 Calculation of Minimum Fluidizing Velocity by different equations

The density of ambient air is obtained from (21)

Temp. ($^\circ$C)	Viscosity (cp)	Kinematic Viscosity $\nu \times 10^{-2}$ ($cm^2\ sec^{-1}$)	Density ($10^{-3} gm/cm^3$)
20	.01813	15.05	1.20465
40	.01908	16.92	1.12765

$\therefore$ At T = 30°c $\qquad \rho = 10^{-3}(1.20465+1.12765)\,\frac{1}{2} = 1.16615\times10^{-3} gm/cm^3$

The gasoline amount is varied in the experiment from 20–107 gm/min it effected very little on the overall density of gasoline-air mixture that fluidized the bed, so an averaged amount will be counted in determining the density of fluidizing medium, gasoline-air mixture. The density of the mixture is varied by the amount of fluidizing air for different particle sizes. The calculation is as the following:-

Average varied gasoline = $(20+34+54+67+80+94+107)\,\frac{1}{7}$ = 65.14 gm/min.

Fluidizing bed area = 100×5 = 500 cm^2

d_p (cm)	U_o^* (cm/sec.)	Rate of air flow $A\,U_o\rho$ (gm/min.)	Av.Rate of Gasoline Mixed (gm/min.)	Total weight of mixture (gm/min.)	Mixture density ($10^{-3} gm/cm^3$)
0.33	52.4	1833.18		1898.32	1.207
0.51	66.8	2336.96	65.14	2402.10	1.199
0.64	105.7	3697.86		3763.00	1.187

*From Table 4.2

From equation 2.12

$$\frac{1.75}{\phi_s\,\epsilon_{mf}^3}\left[\frac{d_p\,U_{mf}\,\rho_g}{\mu}\right]^2 + \frac{150\,(1-\epsilon_{mf})}{\phi_s^2\,\epsilon_{mf}^3}\left[\frac{d_p U_{mf}\,\rho_g}{\mu}\right] = \frac{d_p^3\,\rho_g(\rho_s-\rho_g)g}{\mu^2}$$

Basis PFD size 0.33 cm.

$$\frac{1.75}{(1)(0.401)}\left[\frac{(0.33U_{mf}\ 0.001207)^2}{(0.000186)^2}\right]+\frac{150(1-0.401)}{1^2(.401)^3}\left[\frac{0.33U_{mf}(0.0001207)}{0.000186}\right]$$

$$=\frac{(.33)^3(.001204)(0.02-0.0001207)980}{(.000186)^2}$$

$$182.86\ (0.33)^2\ U_{mf}^2 + 9019.83\ U_{mf}\ (0.33) = (0.33)^3(641050.3)$$

$$182.86\ U_{mf}^2 + 27332.82\ U_{mf} = 211546.5$$

$$U_{mf}^{(1)} = 7.37 \text{ cm/sec.}$$

Similarly for PFD size 0.51 cm. $U_{mf}^{(1)}$ = 9.82 cm/sec.

and for PFD size 0.64 cm. $U_{mf}^{(1)}$ =12.38 cm/sec.

From equation 2.14

$$U_{mf} = \frac{d_p^2\ (\rho_s - \rho_g)\ g}{1650\,\mu} \qquad \text{for } Re < 20$$

Basis PFD size 0.33 cm.

$$U_{mf}^{(2)} = \frac{(0.33)^2(0.02-0.001207)980}{1650\ (0.000186)} = 6.54 \text{ cm/sec}$$

Check Re, $$Re = \frac{(0.33)(6.54)(0.001207)}{.000186} = 13.97$$

---

(1) Calculated from equation 2.12

(2) Calculated from equation 2.14

(3) Calculated from equation 2.15

Similarly for PFD size 0.51 cm, $U_{mf}^{(2)}$ = 15.61 cm/sec. and Re= 51.53
and for PFD size 0.64 cm, $U_{mf}^{(2)}$ = 24.59 cm/sec. and Re=101.87

$\therefore$ for the last two particle sizes, Re do not satisfy the condition in eq. 2.14 that Re<20

From equation 2.15

$$U_{mf}^2 = \frac{d_p(\rho_s - \rho_g)\, g}{24.5\ \rho_g} \qquad Re > 1000$$

Basis PFD size 0.33

$$U_{mf}^{(3)} = \sqrt{\frac{(0.33)(0.018793)980}{(24.5)(0.001207)}} = 14.36$$

Check Re, $Re = \frac{(0.33)(14.36)(.001207)}{.000186} = 30.68$

$\therefore$ U_{mf} = 14.36 was not fitted since Re < 1000

Similarly for PFD size 0.51 cm., $U_{mf}^{(3)}$ = 17.85 cm/sec. and Re= 58.92
and for PFD size 0.64 cm, $U_{mf}^{(3)}$ = 19.9 cm/sec. and Re = 82.82

$\therefore$ for every particle size, Re does not satisfy the condition in eq. 2.12 that Re > 1000

A.2 Calculation of Terminal Fluidizing Air Velocity

From $$C_d Re_p^2 = \frac{4g\, d_p^3\, \rho_g(\rho_s - \rho_g)}{3\mu^2}$$

Basis PFD size 0.33 cm.(range 0.25-0.41 cm)

$$C_d Re_p^2 = \frac{4(980)(.25)^3(.001207)(.018793)}{3(.000186)^2} = 14582$$

from fig 2.12 Re_p = 120

$$\therefore\ U_t = \frac{120(.000186)}{(0.25)(.001207)} = 74.2 \text{ cm/sec.}$$

Similarly for PFD size 0.51 cm, U_t = 114.0 cm/sec
and for PFD size 0.64 cm, U_t = 169.9 cm/sec

A.3 Fluidizing Air Velocities

A.3.1 Fluidizing Air Velocities in the experiment

d_p (cm)	Minimum Velocity U_{mf} (cm/sec)	Superficial Velocity U_o (cm/sec)	Terminal Velocity U_t (cm/sec)
0.33	8.29	52.4	74.2
0.51	8.64	66.8	114.0
0.64	9.66	105.7	169.9

A.3.2 Best Equation determination for the fitness of minimum fluidizing air velocity by sum of square error

d_p (cm.)	Exp. U_{mf} (cm/sec)	$U_{mf}^{(1)}$ (cm/sec)	e= $U_{mf}-U_{mf}^{(1)}$	$U_{mf}^{(2)}$ (cm/sec)	e= $U_{mf}-U_{mf}^{(2)}$	$U_{mf}^{(3)}$ (cm/sec)	e= $U_{mf}-U_{mf}^{(3)}$
0.33	8.29	7.37	.92	6.54	1.75	14.36	6.07
0.51	8.64	9.82	1.18	15.16	6.97	17.85	9.21
0.64	9.66	12.38	2.72	24.59	14.93	17.90	10.24
Sum square of error			9.64		274.55		226.53

A.4 Mode of fluidization

$Fr_{mf} < 1$ shows smooth or particulate fluidization

$Fr_{mf} > 1$ shows bubbling or aggregative fluidization

$$(Fr_{mf})(Re_{p,mf})\left(\frac{\rho_s - \rho_g}{\rho_g}\right)\left(\frac{L_{mf}}{d_t}\right) < 100 \text{ confirms smooth fluidization}$$

$$(Fr_{mf})(Re_{p,mf})\left(\frac{\rho_s - \rho_g}{\rho_g}\right)\left(\frac{L_{mf}}{d_t}\right) > 100 \text{ confirms bubbling fluidization}$$

d_p (cm.)	U_{mf} (cm/sec)	Fr_{mf} $\left(= \frac{U_{mf}^2}{d_p g}\right)$	$Re_{p,mf}$	$\frac{\rho_s - \rho_g}{\rho_g}$	$\frac{L_{mf}}{d_t}$	Index	Mode
0.33	8.29	.213	17.71	15.67	.263	15.49	Smooth
0.51	8.64	.149	28.25	15.68	.263	17.45	Smooth
0.64	9.66	.149	40.02	15.85	.263	24.48	Smooth

Appendix B

B.1 Retention Time of PFD in the Fluidizing Bed.

for d_p = 0.33 cm, the feed rate = 7.8 gm/min.

The bed weight of PFD = 5x100x2.5x0.0116 = 14.5 gm

$\therefore$ Retention Time of PFD in bed is 14.5/7.8 = 1.86 min

Similarly, the other values are obtained as shown below.

d_p (cm)	Fixed Bed Weight W (gm)	Feed Rate F (gm/min)	Ratention Time t (min)	Av.Weight of 100 drops (10^{-3} gm)
0.33	14.5	I = 7.8	1.86	90
		II= 11.3	1.28	
0.51	15.4	I = 6.6	2.33	300
		II= 10.7	1.44	
0.64	15.5	I = 6.3	2.46	350
		II= 10.5	1.48	

B.2 Agglomeration Efficiency

$$\%\text{Agglomeration Efficiency} = \frac{\dfrac{\text{gm.agglomerate}}{\text{gm.gasoline used}}}{\left(\dfrac{\text{gm.PFD}}{\text{gm.gasoline}}\right)_{\sigma = 28\ \text{dyne/cm.}}} \times 100$$

From table 4.6 anexample of calculation will be drawn out by using the data in the group of C average = 150 to present the term %Agglomeration Efficiency and the term %Agglomerate per Feed Rate. All the results have been shown in Table 5.2.

From Table 4.6

Gasoline ml/5min.	Agglomerate							Unagglomerzte	
	Number of drops in an agglomerate cluster						Total Cluster	Weight (gm)	Volume (ml)
	2	(3-4)	(5-6)	(7-9)	(10-20)	>20			
150	4	3					7	39	3500
160	5	8	2				15	39	3500
150	7	2					9	39	3500
150	5	1	2				8	39	3500
155	3	4	-	2			9	39	3500
165	7	1					8	39	3500
150	5	3	3	2			13	38	3400
150	8	1	1	2			12	38	3400
150	7	1	-	1			9	38	3400
150	5						5	39	3500
150	8	3	-	2 1	1		14	39	3500
160	3	6	3	1	1		14	40	3500
155	4	9	2				15	42	3600
160	17	2	-	-	1		20	39	3500
TOTAL 2155	88	44	13	10	3		158	547	

All PFD agglomerated drops $= 88 \times 2 + 44 \times 3.5 + 13 \times 5.5 + 10 \times 8 + 3 \times 15 + 0 \times 20$

$= 527$ drops

Agglomerate weight $= 527 \times 90 \times 10^{-3}/100 \quad = 0.474$ gm.

$$\frac{\text{Agglomerate weight}}{\text{gm. Gasoline used}} = \frac{0.474}{2155 \times 0.68} = 0.000324$$

$$\%\ \text{Agglomeration Efficiency} = \frac{0.000324}{0.2647} \times 100 = 0.1224\ \%$$

$$\%\text{gm. Agglomerate per Feed Rate} = \frac{0.474}{547.474} \times 100$$

$$= 0.08658\%$$

Similarly, in the other group of C, the figures have been found out and shown in Table 5.2.

Appendix C

C.1 Determination of voidage in the fluidizing bed

$$\frac{L_f}{L_{mf}} = \frac{1-\epsilon_{mf}}{1-\epsilon_f}$$

$$\epsilon_f = 1 - \frac{L_{mf}}{L_f}(1-\epsilon_{mf})$$

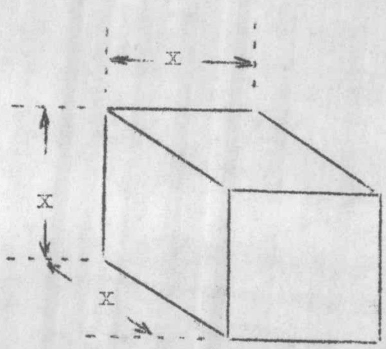
for $\frac{Lmf.}{L_f} = \frac{2.5}{5} = \frac{1}{2}$; $\epsilon_f = \frac{1+\epsilon_{mf}}{2}$

d_p (cm.)	ϵ_{mf}	ϵ_f
0.33	0.401	0.701
0.51	0.385	0.693
0.64	0.382	0.691

C.2 Particle distance in Fluidized Bed

Assume that the PFD in fixed bed was in orthorhombic arrangement. Hence the voidage in the fixed bed was the volume of lattice subtracted by the volume of particle which in this arrangement the total volume of particle in the lattice is equivalent to 2 paticles

Let x was the lattice dimension, it is found that



$$x^3 - \frac{2\pi d^3}{6} = \epsilon x^3$$

$$x^3 = \frac{\pi}{3(1-\epsilon)}(d^3)$$

By the above equation, it would be applied to find the lattice space of particles in fluidizing state

For example

Basis $\qquad d_p = 0.33$ cm, $\epsilon_f = 0.701$ (from C.1)

$$\therefore x^3 - \frac{2\pi(0.33)^3}{6} = 0.701x^3$$

$$(0.299)x^3 = \frac{\pi}{3}(0.33)^3$$

$$x = (1.519)(0.33) = 0.501 \text{ cm}$$

$\therefore$ In fluidizing state, the lattice space would be about 0.501 cm with the volume of 2 molecules occupied

Similarly the lattice space for $d_p = 0.51$ cm, $x = 0.768$ cm.

$d_p = 0.64$ cm, $x = 0.961$ cm.

C.3 Mean Free Path of PFD (20) and number of collisions

$$\lambda = \frac{1}{\sqrt{2}(n\,A)}$$

λ = mean free path (average distance which a molecule travels before it collides with some other molecule in the gas)

n= number of molecules per unit volume

A= total scattering cross section for molecule-molecule collisions.

= πd^2 (if the molecules are indenticle.)

$$\lambda = \frac{1}{\sqrt{2}\,(n\pi d^2)}$$

Example

Basis $\qquad d_p = 0.33$ cm.

from C2, $\qquad n = \dfrac{2}{(0.501)^3}$

$$= \frac{1}{\sqrt{2}\left[\frac{2}{(0.501)^3}\right]\pi(0.33)^2} = .1299 \text{ cm}$$

In addition, to find the number of collision, it would be based on the maximum or terminal fluidizing velocity of particles since the superficial fluidizing velocity of particles was not be able to find out,

$$\text{Max. Number of collision } (N_c) = \frac{\text{Velocity of particle}}{\text{Mean free path}} = \frac{74.15}{.1299}$$

$$= 570.8 \ \frac{\text{collision}}{\text{sec}}$$

d_p (cm)	ϵ_f	x (cm)	λ (cm)	N_c (collisions/sec)
.33	.701	.501	.1299	570.8
.51	.693	.768	.196	581.7
.64	.691	.961	.244	696.4

C.4 Force acting on particles

Total force F_t of the fluid on th sphere is given by the sum of buoyant foce, form drag and friction drag or tangential force

$$F_t = \frac{4}{3}\pi R^3 \rho g + 2\pi\mu Rv + 4\pi\mu Rv$$
$$= \frac{4}{3}\pi R^3 \rho g + 6\pi\mu R v$$

The term $\frac{4}{3}\pi R^3 \rho g$ or buoyant force may be disignated as F_s (the force exerted even if the fluid is stationary) and the term $6\pi\mu Rv$ as F_k (the forcė associated with the fluid movement ie. the kinetic contribution)

For example

Basis $\quad d_p = 0.33$ cm, $U_o = 52.4$ cm/sec

$$F_t = \frac{4}{3}\pi(.33)^3(.001207)980+6 \ (.00186)(.33)(52.4)$$
$$= .1777 + .0607$$
$$= .2384 \text{ dyne}$$

$\therefore$ average force or pressure of fluid distributed on a sphere = $\frac{F_t}{A}$

$$P = \frac{F_t}{A} = \frac{.2384}{(.33)^2} = .6965 \text{ dyne /cm}^2$$

Similar calculation was applied to the other particle sizes and it was observed as illustrated.

d_p (cm)	F_s $\frac{\text{gm cm}}{\text{sec}^2}$	F_k $\frac{\text{gm cm}}{\text{sec}^2}$	F_t $\frac{\text{gm cm}}{\text{sec}^2}$	A cm^2	P dyne/cm^2
0.33	.1777	.0607	.2384	.3423	.6965
0.51	.6599	.1195	.7754	.8175	.9485
0.64	1.296	.2375	1.533	1.287	1.191

C.5 Tensile Strength of Pendular State Agglomerate.

From equation 2.6

$$i_p = 2.78 \frac{(1-\epsilon)}{\epsilon} \frac{\sigma}{d}$$

d_p (cm)	$\epsilon = \epsilon_{mf}$	i_p (dyne/cm^2)
0.33	.401	$\frac{(2.78)(0.401)(28)}{(0.599)(0.33)} = 158$
0.51	.385	$\frac{(2.78)(0.385)(28)}{(0.615)(0.51)} = 96$
0.64	.382	$\frac{(2.78)(0.382)(28)}{(0.618)(0.64)} = 75$

C.6 Tensile strength of Capillary state agglomerate

From equation 2.9 $i_c = \left[\frac{6}{d}\left(\frac{1-\epsilon}{\epsilon}\right) + \frac{c}{A}\right] \sigma \cos.\delta$

Assume that the point of contact is as large as the particle cross-sectional area, so, C is the circum-ference of a particle and A is the cross-sectional area.

$$\therefore i_c = \left[\frac{6(1-\epsilon)}{d\epsilon} + \frac{4\pi d}{\pi d^2}\right]\sigma \cos.\delta$$

$$= (6 - 2\epsilon)\frac{\sigma}{d\epsilon} \qquad (\cos.\delta = 1 \text{ if wet})$$

d_p (cm)	$\epsilon = \epsilon_{mf}$	i_c (dyne/cm^2)
0.33	.401	$6-2(.401)\dfrac{28}{(.33)(.401)} = 1099$
0.51	.385	$6-2(.385)\dfrac{28}{(.51)(.385)} = 746$
0.64	.382	$6-2(.382)\dfrac{28}{(.64)(.382)} = 600$

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